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Comptes Rendus
Proceedings

7e Symposium sur le traitement des eaux usées
7th Symposium on wastewater treatment

Environnement Canada
Environment Canada
AQTE

PROCEEDINGS

7th Symposium on
Wastewater Treatment

November 20-21, 1984
Hôtel Méridien
Montréal

COMPTES RENDUS

7e Symposium sur le traitement
des eaux usées

20 et 21 novembre 1984
Hôtel Méridien
Montréal

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Cat. No. En 44-14/1984
ISBN 0-662-53385-2
BEAUREGARD PRESS LIMITED

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N° de cat. En 44-14/1984
ISBN 0-662-53385-2
IMPRIMERIE BEAUREGARD LIMITÉE

PREFACE

The organizing committee of the 7th Symposium on Wastewater Treatment is extremely pleased to welcome you to this annual event. The constantly increasing success of this conference is the result of your continued support over the last seven years, and it has become a leading forum for the exchange of wastewater technology in Canada.

In addition, the degree of excellence in technology transfer that the symposium has provided has gained international recognition.

The symposium is intended for scientists, engineers, technicians, plant operators and specialists working in research, and the design and the operation of wastewater treatment facilities. The principal objective of this symposium is to promote the exchange of state-of-the-art expertise in the field of municipal and industrial wastewater treatment.

Guest speakers from a variety of specialties and countries will share their most recent experiences and improvements in the wastewater treatment field.

Alain Jolicoeur
Ronald Zaloum
Lisette Provencher

PRÉFACE

Le comité d'organisation du 7ième Symposium sur le traitement des eaux usées est très heureux de vous souhaiter la plus cordiale des bienvenues à cet évènement annuel. Le succès toujours croissant de cette conférence est le résultat de votre support constant durant les sept dernières années et qui a permis d'identifier le symposium comme étant la fine pointe du transfert technologique en matière de traitement des eaux usées au Canada.

De plus, le symposium a acquis une excellente réputation internationale grâce au degré supérieur d'échange scientifique qu'il a réussi à procurer.

Le symposium s'adresse aux chercheurs, ingénieurs, opérateurs, technologues et, donc, aux spécialistes dans les domaines de la recherche, de la conception, de l'opération et de la gestion d'installations de traitement des eaux usées. Le principal objectif du symposium est de promouvoir et de diffuser l'expertise des techniques du traitement des eaux usées domestiques et/ou industrielles.

Des spécialistes de plusieurs pays communiqueront le fruit de leurs expériences, de leurs connaissances et de leurs recherches dans le domaine du traitement des eaux usées.

Jacques Simon
Claude Vergès

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ÉVOLUTION DES AUTOMATISMES DANS LES USINES DE TRAITEMENT DES EAUX RÉSIDUAIRES

Claude Daudenarde
Degrémont Infilco Ltée

RÉSUMÉ

Les usines d'épuration d'eaux résiduaires urbaines, utilisant un même procédé d'épuration biologique, sont d'importance très variable, pouvant traiter les eaux usées d'une population de moins de 5 000 usagers à plus de 1 000 000 d'usagers; les méthodes d'opération et les conceptions des automatismes sont inévitablement différentes entre ces deux extrèmes. Les petites stations sont la plupart du temps exploitées manuellement avec quelques automatismes séquentiels simples et éventuellement un peut de télémétrie. Les grandes usines, pour lesquelles on recherche une optimisation des coûts d'exploitation, une efficacité accrue des procédés, une amélioration des conditions de travail des opérateurs, peuvent justifier beaucoup plus facilement une automatisation poussée. Le système d'automatisme assure alors les deux grandes fonctions distinctes de commande et régulation d'une part et de gestion d'autre part. Le développement très rapide de la micro informatique a facilité grandement l'implantation des automatismes et permet de réduire la taille des usines pour lesquelles ces systèmes s'avèrent rentables. L'avènement des microprocesseurs et le développement des réseaux de communication ont permis d'éliminer les gros systèmes centralisés qui étaient une limite à la fiabilité des automatismes il y a quelques années; on leur préfère maintenant les systèmes décentralisés constitués d'automates locaux répartis par zones géographiques ou fonctionnelles, supervisés par un équipement central chargé de l'acquisition de données, génération de point de consignes, supervision et télédiagnostic, et gestion de l'usine. Les avantages de tels systèmes sont évidents:

- en cas de panne du système central, les équipements continuent de fonctionner en mode automatique
- en cas de panne d'un automate local, seule une partie de l'usine doit passer en mode manuel
- les coûts de raccordements électriques sont considérablement réduits
- meilleur respect des garanties; dans le cas où les procédés mis en oeuvre proviennent de différents fournisseurs, il est important que chaque fournisseur soit responsable de l'automatisation de son procédé; moyennant une bonne définition des besoins de communication, il est alors préférable que chaque équipement soit fourni avec son propre ensemble d'automatisme

L'efficacité, la rentabilité des systèmes et l'obtention des objectifs sont astreintes au contraintes suivantes:

- conception d'automatismes par des automaticiens impliqués dans le procédé
- fiabilité des capteurs
- qualification du personnel d'exploitation; si l'automatisme peut diminuer le nombre de tâches peu valorisantes, en contre partie, si on désire qu'il atteigne son but, la maintenance exige des instrumentistes qualifiés et la conduite de l'usine des décisions sur le procédé

1. RAPPELS SUR L'ÉPURATION DES EAUX RÉSIDUAIRES

L'objectif du traitement des eaux résiduaires est de limiter la pollution déchargée en rivière à une valeur telle que le pouvoir auto-épurateur de cette dernière lui permette de l'accepter sans nuisances. Les eaux usées d'origine domestique contiennent, en général, une faible quantité de matières polluantes (moins de 1 g/L) mais les volumes d'eau à rejeter sont considérables (environ 400 L/jour/habitant) et motivent leur traitement. L'épuration de l'eau fait, en général, appel à des processus physiques de décantation, pour la rétention des matières en suspension et biologiques pour l'élimination des matières dissoutes et colloidales. Dans le processus biologique, l'épuration s'effectue grâce à des micro-organismes qui utilisent les matières organiques contenues dans l'eau pour leur croissance et leur multiplication. Le procédé généralement utilisé, dit à boues activées, est aérobie et nécessite un apport d'oxygène, gros consommateur d'énergie.

Les boues provenant du traitement constituent un produit extrêmement putrescible qu'il convient de rendre inoffensif avant de l'éliminer. Leur part organique relativement élevée (de l'ordre de 60 à 75%) peut être plus ou moins convertie en énergie par des procédés tels qu'une fermentation anaérobie, productrice de gaz méthane ou une incinération des boues dont la teneur en matières solides aura été portée préalablement à une valeur suffisante par une filtration précédée d'un conditionnement.

2. L'USINE D'ÉPURATION

Les installations de traitement d'eaux résiduaires domestiques utilisant le procédé biologique à boues activées couvrent une gamme d'usagers très large, qui peut aller de quelques centaines d'usagers à plusieurs millions; le nombre et l'importance des ouvrages de traitements et équipements annexes peuvent donc être très variables mais le principe du traitement demeure. Toutes comportent au moins un bassin d'aération et un clarificateur destinés à séparer la boue activée de l'eau interstitielle épurée, pour la recycler et maintenir une concentration suffisante du milieu en micro-organismes.

Une usine typique moyenne (50 000 usagers, par exemple) comporterait:

- pour le traitement de l'eau:
  - Un pompage en tête de relèvement des eaux brutes
  - Un ensemble d'ouvrages de prétraitement, avec dégrillage, dessablage, dégraissage
  - Une décantation primaire avec extraction des boues par pompage vers le traitement des boues
  - Un traitement biologique avec bassins d'aération et clarificateur de séparation. Les deux systèmes d'aération les plus courants sont soit l'insufflation d'air dans le milieu par des soufflantes extérieures au moyen de diffuseurs immergés, soit l'aération mécanique au moyen d'aérateurs de surface (ou quelquefois immergés) locaux
  - Un recyclage partiel des boues séparées, l'excès de boues correspondant à la prolifération bactérienne étant retourné vers le décanteur primaire
  - Éventuellement un traitement de finition

- pour le traitement des boues:

  - Un épaississement des boues fraîches
  - Une digestion anaérobie (minéralisation de la boue avec production de méthane) chauffée et brassée
  - Un conditionnement chimique ou thermique de la boue digérée
  - Une déshydratation mécanique sur filtre-presse

Dans tout ce qui précède, l'aération est, de loin, le plus gros consommateur d'énergie et c'est sur ce point qu'un automatisme bien adapté peut être le plus rentable.

3. LA CONSOMMATION D'ÉNERGIE

On peut répartir la consommation d'énergie d'une usine d'épuration en trois catégories principales:

- l'énergie électrique reliée aux moteurs pour l'aération des boues, les pompes, râcleurs des décanteurs, etc., ainsi que l'éclairage des bâtiments
- l'énergie thermique pour le chauffage des digesteurs de boues, le conditionnement thermique des boues, le chauffage des bâtiments
- les diverses énergies reliées à la préparation des produits chimiques éventuellement utilisés

Le tableau no 1 donne à titre d'exemple la consommation d'électricité en Wh/m^3 d'eau pour les postes principaux d'une station typique fonctionnant à sa charge normale et recevant des eaux de provenance urbaine; les trois colonnes correspondent à des valeurs moyennes pour trois différentes tailles d'usines exprimées en milliers d'usagers (10, 100, 200 milliers d'usagers) et sont basées sur un relevage en tête de station de 5 mètres.

	Milliers d'usagers		
	10	100	200
Relevage EB	20	15	15
Prétraitement (Dégrill.Déssabl.Dégraiss.)	20	15	15
Décantation primaire et pompage BF	5	3	3
Épuration biologique			
- Aération	310	220	200
- Clarification/Recirculation	25	20	20
Traitement des boues			
- Épaississement	-	5	5
- Digestion anaérobie	30	20	20
- Déshydratation mécanique	25	10	10
Total en W.h/m3 d'eau	435	308	288

- Soit, rapporté à la DBO éliminée, en considérant 225 mg DBO_5/m^3			
kW.h/kg DOB_5	1,94	1.37	1.28
- Soit, rapporté à l'usager en considérant un volume d'eau de 250 L/jour/usager			
en W.h/jour.usager	109	77	72

CONSOMMATIONS MOYENNES D'ÉNERGIE ÉLECTRIQUE DANS UNE STATION D'ÉPURATION D'EAUX RÉSIDUAIRES URBAINES

TABLEAU 1

Les valeurs indiquées ne tiennent pas compte de l'énergie de chauffage, d'éclairage ainsi que des possibilités de récupérer de l'énergie. Comme on peut le constater, les consommations d'énergie par usager sont inversement proportionnelles à la taille de l'usine. Si on se place du point de vue de l'économie d'énergie, on a donc intérêt à regrouper le plus grand nombre possible d'usagers de façon à construire des usines aussi grosses que possible d'autant plus que c'est dans de grosses usines qu'on peut le plus facilement implanter des systèmes automatiques qui en optimisent encore plus le fonctionnement. Toutefois, il y a d'autres paramètres à considérer comme, par exemple, le coût des réseaux d'égouts, d'autant plus que la consommation d'énergie ne représente qu'une partie des dépenses d'exploitation (10 à 30%); d'autres postes, comme l'amortissement du capital ou les coûts de personnels, peuvent modifier cette approche. Dans tous les cas, la consommation d'énergie sera optimisée si l'usine est bien adaptée aux besoins, si elle est opérée dans des conditions proches des valeurs de design et si elle est correctement conduite et entretenue.

4. INSTRUMENTATION ET AUTOMATISMES

Pendant longtemps, les usines d'épuration d'eaux résiduaires ont été le "parent pauvre" de l'automatisme; alors que tant d'industries en étaient déjà à des degrés d'automation élevés, la plupart des stations de traitement d'eaux usées de par le monde étaient encore opérées "à la mitaine".

Il y a à cela plusieurs raisons:

- l'eau résiduaire n'est pas un produit commercialisable et il semblait peu rentable d'utiliser des techniques de pointe pour aider à traiter un produit destiné à la rivière
- le produit à traiter est très variable en quantité (grosses variations de débit) et en qualité (charge de pollution, produits toxiques, détergents, etc.); tous ces paramètres sont bien plus difficiles à contrôler que les paramètres physiques habituels tels que niveaux, débits, températures, etc., et les algorithmes connus (régulation proportionnelle, intégrale et dérivée essentiellement) sont inadaptés en raison surtout des long temps de réponse des ouvrages.

Avant de penser à implanter des systèmes d'automatismes, il a donc fallu:

- définir les paramètres importants devant entrer dans une boucle de régulation
- inventer ou améliorer les capteurs pour mesurer ces paramètres
- définir les modèles mathématiques servant à l'automatisation des procédés et trouver les régulateurs adaptés à l'exécution de l'algorithme

Dans le même temps, on prenait conscience des implications sur l'environnement d'un mauvais traitement des eaux résiduaires et qu'une automatisation des procédés s'imposait. Dans la plupart des pays industrialisés, l'implantation d'usines de traitement des eaux résiduaires s'est fait en même temps que le développement des systèmes d'automatisation informatiques, systèmes qui étaient les mieux adaptés à l'élaboration et à l'application de modèles mathématiques plus ou moins complets et sujets à évolutions. C'est donc vers de tels systèmes qu'on s'est très vite tourné. Les premiers systèmes utilisés, comme dans toutes les autres industries, étaient de gros systèmes centralisés avec des logiciels très longs à mettre au point et une fiabilité discutable puisque tout l'automatisme de l'usine était assujetti à une seule unité de calcul; en cas de panne de cette unité, c'est toute l'usine qui devait être opérée en mode manuel. Avec le développement ultra-rapide de la micro-informatique, les systèmes "bas de gamme" sont devenus de plus en plus accessibles et on préfère maintenant s'orienter vers des systèmes décentralisés, le système central ayant plutôt un rôle de supervision et de gestion. La popularité et le faible coût des petits systèmes permet également d'envisager le développement des automatismes sur les petites usines. Il reste que l'implantation d'automatismes continue de poser certains problèmes importants en raison principalement:

- D'une approche généralement trop compartimentée au niveau de la conception initiale; il faut absolument que l'automaticien soit impliqué de très près dans le fonctionnement du procédé et possède l'expérience des problèmes reliés aux différentes mesures
- Du manque de capteurs ou du manque de fiabilité de certains capteurs; on sait, par exemple, que la DBO5 est une des mesures les plus représentatives de la pollution d'une eau, mais comment baser un automatisme sur une mesure qui prend 5 jours?
- De la formation du personnel d'exploitation; il y a encore trop de gens qui pensent qu'une automatisation poussée, élimine la nécessité de personnel compétent. Cette approche est la cause de beaucoup de déboires car bon nombre d'usines automatiques ne fonctionnent pas correctement en raison de défaillance dans l'entretien des instruments de contrôle, due au manque d'instrumentistes, ou d'une connaissance insuffisante du procédé

En fonction de l'importance de l'usine, on peut dégager les tendances suivantes:

Petites usines pour moins de 10 000 usagers: Généralement pas de personnel à demeure mais des visites journalières d'entretien et de relevés. Généralement, on préfère surdimensionner les ouvrages pour obtenir la sécurité de traitement. L'instrumentation se réduit le plus souvent à une mesure de débit et éventuellement de l'oxygène dissous. Le reste des automatismes est principalement constitué par des opérations séquencielles ou combinatoires: arrêt, départ de pompes, dosage de produits chimiques par minuteries, extraction ou recirculation de boues sur base de temps ... Ces opérations sont généralement exécutées avec succès par des automates programmables. Les paramètres principaux sont souvent retransmis à l'hôtel de ville par un système simple de télémétrie permettant d'avertir le responsable en cas de défaillance majeure.

Usines de moyenne importance (20 000 à 100 000 usagers): Les coûts d'investissement et d'exploitation commencent à devenir importants et justifient l'implantation d'automatismes plus poussés permettant d'optimiser le dimensionnement des installations et leur

rendement énergétique; on peut, par exemple, envisager l'utilisation de micro-ordinateurs pour la régulation d'oxygène dissous incluant l'optimisation de la marche des soufflantes, plusieurs automates programmables gérant, par équipement ou par groupes d'équipements les informations logiques ... etc. Dans ce type d'usine, il y généralement du personnel à demeure mais la taille de l'usine ne justifie pas une équipe de spécialistes très importante; on cherchera donc à limiter la maintenance en utilisant l'instrumentation la plus éprouvée et conventionnelle.

Grandes usines (supérieure à 200 000 usagers): l'optimisation des coûts de mise en oeuvre et de fonctionnement, l'importance des résultats d'épuration et leur impact sur l'environnement justifient une instrumentation et une automatisation poussées. La présence, sur place, d'une équipe d'instrumentistes permet l'utilisation de capteurs plus élaborés mais demandant plus d'entretien; on peut aussi penser à des systèmes de régulation évolutifs ou le système informatique permet de développer et d'améliorer, avec l'accumulation de données, le modèle mathématique de régulation de l'usine.

Quelle que soit la dimension de l'usine, il y a toujours des règles élémentaires à respecter:

- Un automatisme, aussi complexe soit-il, ne peut pas remplacer un équipement mal dimensionné ou mal adapté; il permet simplement d'optimiser au maximum son dimensionnement et son fonctionnement, simplifier et faciliter l'exploitation en mécanisant les tâches déplaisantes ou dangereuses, assurer une meilleure protection des personnes et des équipements.

- L'automatisme fait autant partie d'un équipement que le moteur fait partie d'une auto; l'un et l'autre sont indissociables. Le bon fonctionnement d'un automatisme est donc relié à une parfaite connaissance et une bonne expérience du procédé qu'il contrôle; réciproquement, le calcul d'un procédé pourra varier selon le type d'automatisme dont il sera équipé.

- Un système automatique peut tomber en panne et il est essentiel dans la conception du système de commandes d'en tenir compte; il y a donc lieu de prévoir avec autant de soin que pour la marche automatique les conditions de marche manuelle. Le système de secours manuel doit continuer d'assurer la sécurité des personnes et du matériel; il doit être fonctionnel en n'obligeant pas les opérateurs à courir d'un bout à l'autre de l'usine. Par contre, en mode manuel, l'opérateur sera appelé à prendre des décisions normalement confiées au système de commandes et c'est là que sa qualification aura toute sa signification. Des décisions erronées peuvent entraîner des conséquences graves: détérioration du traitement, bris de matériel, surconsommation d'énergie ...

5. EXEMPLES D'APPLICATIONS

5.1 L'aération des boues activées

Parmi les régulations qu'on rencontre le plus souvent, la régulation de l'aération des boues activées est la plus importante. C'est le poste représentant la plus grande consommation d'énergie et, pour beaucoup de petites et moyennes usines, c'est la seule régulation existante.

Les conditions de rentabilité d'une telle régulation sont les suivantes:

- Il faut que les nécessités de brassage ne soient pas prépondérantes sur les besoins biologiques en oxygène.
- Il faut une garantie que les sondes de mesure d'oxygène soient fiables et convenablement entretenues.
- Il faut que la régulation maintienne à son rendement optimal chacun des équipements du système (aérateur, soufflantes, diffuseurs).

Dans ce qu'on a appelé une petite station, la marche des aérateurs est généralement asservie à une horloge journalière programmable. À partir des enregistrements de l'oxygène dissous, l'opérateur peut déduire les intervalles et durée de marche des aérateurs requis en fonction des variations journalières et programmer son horloge en conséquence.

Dans une station de moyenne importance, le système d'aération pourra, par exemple, être contrôlé par un micro-ordinateur (de la taille d'un ordinateur personnel) à partir de la mesure d'oxygène dissous dans le bassin. L'ordinateur pourra optimiser le nombre et le point de fonctionnement des soufflantes, générer des rapports et des alarmes.

Les informations principales à raccorder au système pourront être:

- Entrées de la mesure d'oxygène dissous, du débit de chaque soufflante, de la position des ventelles de modulation, de la position des vannes de distribution d'air, de la pression d'air, des kilowatts consommés par chaque soufflante, des interrupteurs de fin de course de vannes manuelles sur le circuit, des instruments de surveillance de l'équipement comme, par exemple, la température des paliers des surpresseurs, éventuellement du débit d'entrée d'eau au bassin ... etc.
- Sorties vers les vannes modulantes et les ventelles, arrêt-départ des soufflantes, alarmes à distance ... et de préférence un écran cathodique et une imprimante.

Toutefois, l'expérience nous a montré qu'une régulation uniquement basée sur la mesure d'oxygène dissous pouvait présenter des aléas si l'on ne prenait pas en compte les deux paramètres suivants:

- la nécessité de remettre les boues en suspension périodiquement
- la possibilité de nitrification des boues

En effet, il peut arriver que la demande en oxygène de la boue soit telle que les besoins en brassage ne soient plus totalement satisfaits et que les boues aient tendance à décanter; pour remédier à cela, il y a lieu de prévoir, dans le système de régulation, un brassage forcé périodique quelle que soit la teneur en oxygène dissous. Ce brassage forcé est programmé sur horloge, de préférence en fonction des heures de la journée où l'on sait que la charge en pollution peut être la plus forte.

Dans le cas d'aération prolongée, on assiste au phénomène de nitrification (oxydation de l'azote organique et ammoniacal). Si, par exemple, on utilise des aérateurs programmés en "tout ou rien", on constatera, après l'arrêt de l'aérateur, une chute brutale du taux d'oxygène dissous; ce déficit apparent en oxygène favorisera la dénitrification. On réalisera une économie substantielle d'énergie en prenant en compte ce phénomène dans la régulation; le redémarrage des aérateurs devra être temporisé de façon à créer des zones temporaires d'anoxie.

Dans une grande usine, le principe précèdent sera conservé mais, la taille des ouvrages étant plus grande, il faudra généralement un système plus élaboré. Plusieurs

points de mesure par bassin seront souvent nécessaires; l'automate devra recueillir ces valeurs, éliminer les valeurs considérées comme erratiques, faire des moyennes, etc ... De plus, il y a rarement un seul bassin; l'automate devra donc optimiser l'aération sur les différents bassins, tout en maintenant l'énergie globale dépensée à son taux minimum. Comme l'automate local est généralement raccordé à un système central, il dispose alors d'une banque de données et une puissance de calcul considérables qui permettent, dans une certaine mesure, de prévoir certaines perturbations en fonction de variations à un autre point de l'usine; de plus, on peut aussi envisager une régulation auto-adaptative dont les paramètres de régulation et même l'algorithme de régulation évoluent dans le temps en fonction de variations passagères où d'un changement dans la qualité du produit à traiter. On retrouve ici, mais avec encore plus d'acuité, les problèmes communs à toutes les usines de traitement d'eau qui sont reliés aux très grandes constantes de temps des bassins ainsi qu'aux temps de réponse des réactions chimiques et biologiques; le résultat d'une variation du débit d'air se fera sentir après des temps aussi longs que 15 à 20 mn.

5.2 L'extraction et la recirculation des boues biologiques

Dans une station d'épuration fonctionnant à forte ou moyenne charge massique, les problèmes de gestion des boues dans le couple bassin d'aération-clarificateur est plus délicat que dans une usine fonctionnant en aération prolongée. En effet, d'une part l'inertie due à la masse de boues présente dans le bassin est insuffisante pour absorber les variations de débit d'eau usée et il convient de réguler le débit de recyclage des boues activées en conséquence et, d'autre part, l'indice de volume de boues est relativement variable à l'intérieur d'une journée. L'expérience montre que l'on peut automatiser la gestion des boues activées (recyclage et extraction des boues en excès) à partir de capteurs photométriques. L'usine d'épuration de COMPIEGNES (FRANCE) fonctionne depuis plus de 5 ans avec ce type de capteurs et il a été possible d'établir une correlation quasi linéaire entre les signaux délivrés par ces capteurs et la concentration des boues. Un déficit en boues, détecté dans le bassin d'aération, augmente automatiquement le débit de recirculation des boues jusqu'à l'obtention d'une concentration acceptable. Inversément, un excédent de boues, constaté même après réduction de la recirculation provoque l'extraction des boues en excès.

5.3 Automatisation d'un épaississeur gravitaire de boues

Le bon fonctionnement d'un épaississeur de boues nécessite une alimentation en boues fraîches la plus régulière possible ainsi que le maintient d'une tranche d'eau au-dessus des boues très importante, de façon à garantir la qualité de la surverse. L'expérience a montré que l'automatisation de la récupération des boues doit tenir compte des réactions de l'épaississeur aux variations des conditions d'alimentation. Ce type de régulation a été réalisé par DEGRÉMONT à l'usine d'épuration d'ALENÇON (FRANCE) et a donné lieu à une série d'essais sur différents capteurs avant de retenir la configuration suivante:

- sur l'alimentation en boues fraîches:
  - débit: mesuré par débitmètre magnétique (les débitmètres ultrasoniques à effet DOPPLER ont également donné de bons résultats)
  - concentration en MES: turbidimètre optique, équipé d'un dispositif mécanique de nettoyage en continu de la chambre de mesure
  - pH: pHmètre avec système de nettoyage par injection d'eau
- sur l'extraction des boues épaissies:
  - débit: débitmètre magnétique

- concentration en MES: analyseur de concentration de type ultrasonique monté dans la ligne d'extraction
- pH: pHmètre avec nettoyage par injection d'eau

- surversé:

  - concentration en MES: turbidimètre optique à lumières alternées
  - voile de boues: détecteur ultrasonique

Le principe de régulation est le suivant: Le puits d'alimentation en boues est en communication continue avec le décanteur primaire. Le fonctionnement de la pompe d'alimentation est asservi à l'analyseur de concentration des boues fraîches; lorsque celle-ci descend en-dessous d'une limite prédéterminée, la pompe est arrêtée. Elle est redémarrée ensuite, après une temporisation. L'intégration du débit entrant reliée à la mesure de concentration permet de connaître la quantité de MES admise sur l'épaississeur.

La mesure de pH des boues fraîches contrôle l'admission de chaux de façon à maintenir le pH entre 6 et 8.5. L'admission de boues fraîches dans l'épaississeur entraîne la montée du voile de boues; le détecteur de voile de boues démarre alors la pompe d'extraction des boues épaissies qui s'arrête lorsque la concentration de celle-ci descend en-dessous d'une valeur prédéterminée. L'intégration du débit des boues épaissies, reliée à la mesure de concentration en MES permet de doser automatiquement les réactifs de conditionnement en vue d'une déshydratation.

Comme on peut le constater, c'est l'automatisme de l'épaississeur qui gère les extractions du décanteur primaire et, dans une certaine mesure, conditionne l'automatisme du système de déshydratation.

6. LES CAPTEURS

Encore une fois, les capteurs demeurent un obstacle au développement des automatismes en traitement des eaux résiduaires en raison de la nature du produit à analyser et de la complexité et la variété des paramètres physico-chimiques à analyser. Ils sont donc soumis à rude épreuve:

- Bouchages par les matières en suspension (minérales ou organiques, graisses, filasses, celluloses, matières plastiques ...) ou par des proliférations bactériennes.
- Encrassement progressifs ou corrosion amenant une dégradation progressive de la validité des mesures. Certains capteurs, telles les sondes de mesure de pH ou encore les débitmètres magnétiques peuvent être pourvus de nettoyages ultrasoniques automatiques et il ne faudra pas hésiter à les utiliser.
- Présence de produits toxiques ou apport non prévus d'ions gênants perturbant la mesure.
- Incompatibilité entre le type de mesure et les caractéristiques de l'eau à mesurer: par exemple, certaines mesures ne peuvent se faire que par photométrie et peuvent être largement perturbée par la couleur de l'eau intersticielle, ou le taux de matières en suspension ou la taille du floc organique, etc.

On peut mesurer un débit par un débitmètre à ultra-sons mais celui-ci pourra être influencé par le taux de MES; réciproquement, on peut mesurer le taux de matières en suspension par des capteurs ultra-soniques mais ceux-ci peuvent subir une influence du débit.

À titre d'exemple, on peut considérer la mesure d'oxygène dissous qui est une des mesures fondamentales; cette mesure se fait au moyen d'une sonde polarographique dite sonde de CLARK associée à une thermistance de correction de température. Bien que considérée généralement comme fiable, cette mesure peut donner des résultats erronés si on ne sait pas qu'elle peut être influencée par:

- une vitesse de balayage trop faible ou trop variable
- des variations importantes de la pression atmosphérique
- des variations importantes de la température dépassant les possibilités de correction de la thermistance (trop grande incertitude sur la loi de transfert de l'oxygène à travers la membrane de la sonde)
- des variations importantes dans la salinité de l'eau
- la qualité de la membrane
- un vieillissement dû à l'usure par électrolyse de l'anode d'argent
- interférences de produits chimiques
- interférences de bulles d'air

Comme dans le domaine des ordinateurs, l'apparition du microprocesseur a permis de faire un grand pas dans l'électronique associée aux capteurs; en raison du faible coût des microprocesseurs on rencontre de plus en plus de capteurs équipés de tels composants. Le traitement de l'information par des procédés numériques permet d'éliminer presque complètement les dérives dues aux composantes électroniques; les appareils équipés de microprocesseurs possèdent un programme d'auto-diagnostic qui permet d'identifier très vite la cause d'une défaillance. De plus, les microprocesseurs intégrés aux capteurs permettent de délivrer au système de commande une information généralement plus directement exploitable; à ce sujet la mesure de débit sur déversoir ou dans un canal est un exemple typique. Dans un canal de type PARSHALL, le débit est une fonction exposant 3/2 de la hauteur de lame; il y a moins de dix ans la mesure de débit impliquait:

- Mesure de la hauteur de lame par un procédé à flotteur ou de bulle-à-bulle.
- Linéarisation de la courbe exp. 3/2 soit par un procédé mécanique intégré au capteur (came caractérisée) soit par un procédé de calcul; $h^{3/2}$ peut se calculer par le produit de h par la racine carrée de h ce qui impliquait un extracteur de racine carrée et un multiplicateur analogique. Si on remplace le PARSHALL par un déversoir en V la loi de variation devient alors en exposant 5/2 et on doit alors effectuer le produit de h par h par racine carrée de h.

Aujourd'hui la mesure de niveau par ultrason permet d'avoir un capteur sans contact avec le liquide et le microprocesseur intégré à l'appareil de mesure permet de générer un signal de sortie linéaire avec le débit, programmable en fonction du type d'élément primaire utilisé, avec l'avantage supplémentaire qu'on peut utiliser dans toute l'usine un seul type d'instrument quel que soit le déversoir et quelle que soit la gamme de débit à mesurer et donc faciliter la maintenance.

Le choix des capteurs doit donc, au moment de la conception de l'usine, faire l'objet d'une étude poussée; par exemple, on peut en venir à la conclusion qu'il est préférable d'effectuer certaines mesures manuellement, en laboratoire, et entrer manuellement les résultats dans le dispositif de régulation, que de continuellement entretenir un appareil de mesure dont on pourra toujours douter. C'est le cas de mesures telles que la DCO (demande chimique en oxygène), la DTO (demande totale en oxygène), le COT (carbone organique total); les essais d'automatisation d'appareils d'analyse qui donnaient satisfaction en laboratoire ont montré que, en application industrielle, certains de ces appareils demandaient un entretien important.

On pourra aussi remplacer la mesure peu fiable de certains paramètres par d'autres mesures beaucoup plus fiables, portant sur d'autres paramètres mais donnant une bonne approximation; par exemple, la mesure du couple d'entraînement d'un râcleur d'épaississeur peut donner une bonne approximation de la concentration de la boue. On devra également profiter au maximum de l'expérience des fournisseurs dans le domaine; par contre, on devra s'assurer que les références données par le constructeur correspondent bien au même type d'application. À ce sujet, on peut prendre pour exemple le "SÉDIMOMÈTRE" développé par DEGRÉMONT pour la mesure et la transmission du pourcentage de boues dans un bassin d'aération et qui permet d'automatiser l'extraction des boues en excès. Cet appareil donne d'excellents résultats en aération classique mais en aération prolongée, avec le phénomène de nitrification-dénitrification, il se produit dans l'appareil des phénomènes de dégazage qui limitent et rendent peu fiable l'utilisation de ce capteur dans un tel cas.

La conception du système d'automatisme devra également tenir compte des incertitudes mentionnées plus haut; on pourra, par exemple:

- doubler ou tripler les capteurs les plus vitaux, comparer les signaux et éliminer ceux qui paraissent aberrants
- effectuer des moyennes sur des périodes de temps plus ou moins courtes pour éliminer les erreurs transitoires
- éviter que les alarmes ne prennent des décisions sur des mesures erronées
- effectuer des comparaisons entre le signal lu et ce qu'il devrait théoriquement être en fonction du modèle mathématique et des autres mesures et donner une alarme s'il s'écarte notablement de cette valeur théorique

Quand on sait que sur une grande station il est fréquent d'atteindre les 200 points de mesure (en ne considérant que les mesures analogiques les plus classiques), on comprend l'importance qu'il faut attacher à ce problème. Ces mesures se répartissent en:

- mesures de débit: eaux brute, décantée, épurée, boues en excès, boues en retour, boues fraîches, boues épaissies, boues digérées, air des soufflantes, gaz, etc.
- mesures de niveaux: niveaux d'eau dans les différents bassin, niveaux de boues dans les décanteurs, clarificateurs ou épaississeurs
- mesures d'oxygène dissous dans les boues activées
- mesures de concentration en MES dans les boues activées, dans les retours de boues, dans les boues épaissies, dans les boues digérées
- mesures de températures: dans les digesteurs, dans l'air d'aération des boues activées, dans les échangeurs
- mesures de quantité d'énergie dans tous les équipements gros consommateurs d'énergie
- mesures de pression: gaz des digesteurs, gaz de brassage, air des boues activées

7. MICRO-INFORMATIQUE ET STRUCTURES HIÉRARCHISÉES

Comme il en a déjà été fait mention, la particularité des algorithmes de régulation (long temps de réponses, paramètres difficiles à mesurer, nécessité d'accumuler des données) jointe à la nécessité d'avoir des systèmes évolutifs en raison des nombreuses inconnues lors de l'élaboration des procédés, incitent à l'automatisation par des moyens informatiques; dans les années '70, on ne disposait que de gros systèmes très coûteux et il

n'y avait pas d'autre choix que la centralisation avec les inconvénients que nous avons déjà cités. À l'ère de l'ordinateur personnel, tout le monde s'accorde pour préconiser la décentralisation du traitement de l'information en gardant au système central le rôle de surveillance et de gestionnaire. L'exemple précédent de transmetteur de débit à microprocesseur représente un cas de décentralisation à l'extrême puisque l'information est alors traitée au niveau même du capteur. La décentralisation des décisions est toujours accompagnée d'une hiérarchisation des systèmes, c'est-à-dire qu'il n'y a pas nécessairement deux niveaux (un central et un local) mais plusieurs niveaux, un automatisme local pouvant être surveillé par un autre automate local d'un plus haut niveau, etc. Par exemple, la logique de commande d'un poste de pompage dans le bâtiment de traitement des boues peut être une logique cablée puisqu'il s'agit simplement d'agir sur des démarreurs de moteurs à partir de niveaustats; par contre, la surveillance du poste de pompage peut se faire à partir d'un automate supervisant tout le bâtiment de traitement des boues et c'est cet automate qui rend compte au système central.

Les avantages principaux de la décentralisation peuvent être résumés ainsi:

- Élimination de la possibilité d'une panne sur toute l'usine.

- Meilleure répartition des responsabilités; il est toujours préférable de laisser à un fournisseur de procédé la responsabilité de l'automatisation et de l'instrumentation relative au procédé qu'il fournit de façon à éviter les rejets de responsabilité en cas de non respect des garanties. De plus, dans la plupart des cas, les fournisseurs disposent de logiciels déjà éprouvés pour le type d'équipement fourni qui sont aisément adaptables à chaque application et à faibles coûts. Ceci n'élimine pas la nécessité de spécifier dans le cahier des charges les performances du système ainsi que les besoins spécifiques à l'installation tels que le type de dispositif de secours manuel, le type d'interface entre l'automate local et le réseau de communication vers le système central ou d'autres équipements, ainsi que le type et nombre d'informations à transmettre.

- Meilleure adaptation de l'automate au type de procédé qu'il contrôle: l'ordinateur, dans sa définition large est presque universel mais moyennant des coûts de logiciel importants. Par contre, il existe sur le marché bon nombre d'automates adaptés à des tâches spécifiques; par exemple, s'il s'agit d'une tâche de pure logique combinatoire ou séquencielle on utilisera un automate programmable de préférence à tout autre système en raison de la simplicité de sa programmation. D'autres systèmes sont adaptés à l'acquisition et à la retransmission de données, d'autres à la régulation analogique, etc. Partout où on peut remplacer la programmation par de la configuration à partir de logiciels existants, on aura avantage car la mise au point de nouveaux logiciels continue à être coûteuse.

- Réduction considérable des fils de liaison qui sont alors remplacés par un "bus" de communication et de ce fait amélioration de la qualité de l'information provenant des capteurs puisqu'elle est transmise sur une courte distance, tout en diminuant considérablement les coûts de câblage.

Au moment de la conception d'un système décentralisé, il faut:

- Étudier avec beaucoup de soin le "découpage" des équipements de traitement et parfaitement identifier ce qui devra constituer un ensemble fonctionnel. Par exemple, le système d'aération est constitué par des soufflantes, des diffuseurs, des analyseurs d'oxygène dissous, un dispositif de modulation et distribution d'air dans

les bassins, un automate de régulation d'oxygène dissous et du contrôle des soufflantes; cet ensemble ne devrait pas être morcelé et devrait faire l'objet d'une seule fourniture.

- Définir avec précision la hiérarchisation des décisions et le type d'interface entre les différents niveaux.

Les spécialistes s'accordent pour dire que les prochaines années seront celles des réseaux de communication comme les dix dernières années ont été celles des microprocesseurs. En effet si, théoriquement, tous les systèmes informatiques peuvent communiquer entre eux à travers des interfaces standardisées (du type RS 232, RS 422, IEEE 488, par exemple), dans la pratique cela exige souvent le développement de protocoles de communication souvent très délicats et prohibitifs pour une application particulière. Si, d'un côté, on cherche à optimiser le type d'automate en fonction de la tâche qu'il a à accomplir, il ne faut par perdre cet avantage en compliquant le processus de communication. Chaque jour, les manufacturiers et les compagnies de logiciels sortent de nouvelles possibilités de communication entre systèmes provenant de différents manufacturiers et ce phénomène ira en s'accentuant.

Le système central

Celui-ci est normalement lui-même hiérarchisé, c'est-à-dire qu'il n'est plus constitué d'une seule unité centrale mais plutôt d'un ensemble de microcalculateurs, chacun de ces calculateurs ayant une tâche bien définie: gestion des écrans cathodiques, des imprimantes, du stockage des données sur disque dur ou sur disquettes, des claviers, etc. Bien souvent on associera au système de commandes, un système totalement indépendant du premier mais pouvant profiter des données de celui-ci pour effectuer les travaux de calcul scientifique, de gestion de l'usine tel les bilans énergétiques, les consommations de réactifs ou même éventuellement du traitement de texte. Le système central est souvent utilisé pour la surveillance d'installations à l'extérieur de l'usine (comme les réseaux, par exemple).

Le dialogue homme-machine s'effectue à partir des terminaux du système:

- Les écrans cathodiques ayant souvent des tâches définies, comme par exemple un écran réservé à la lecture des alarmes, un autre au synoptique du procédé, un troisième à la lecture des données. La lecture se fait généralement par pages hiérarchisées, certaines pages fournissant des informations très générales sur un grand nombre de points, d'autres au contraire donnant des informations détaillées sur un nombre restreint de points.

- Les imprimantes permettant d'éditer des rapports périodiques ou d'événements, des courbes de relevés.

- Les claviers qu'on devra essayer de rendre aussi accessibles que possible pour tout ce qui est en rapport avec l'exploitation courante.

On peut en effet définir deux niveaux de dialogues entre l'homme et la machine:

- Les dialogues de conduite de procédé qui doivent être accessibles au plus grand nombre possible d'exploitant et qui peuvent impliquer des décisions rapides qui excluent le "pitonnage"; il faut donc que les organes de dialogues soient clairs et les opérations simples.

- Les dialogues de création, accessibles généralement à un personnel plus restreint et plus spécialisé et qui permettent de modifier la configuration du système. La plupart des automates possèdent une protection par clé, rendant ce type de dialogue accessible aux seules personnes autorisées.

Selon le type d'automate, et selon le logiciel fourni par le constructeur, il peut être possible pour le client d'effectuer certaines modifications à la configuration du système sans disposer d'une équipe d'informaticiens chevronnée; là encore, il y aura lieu de juger si cette approche se justifie ou s'il est préférable de négocier avec le constructeur une entente à long terme pour effectuer les modifications éventuelles au système.

8. CONCLUSION

Cet exposé ne prétend pas apporter des solutions aux problèmes posés par l'implantion d'automatismes dans les usines de traitement d'eaux résiduaires mais plutôt effectuer un survol des questions qu'on doit se poser au moment de la concepion d'un système, les réponses étant particulières à chaque application.

Nous avons pu dégager deux idées principales:

- Comme dans n'importe quel autre type d'usine, l'automatisme ne peut être efficace que s'il est conçu par des personnes fortement impliquées dans le procédé.

- Une caractéristique des usines d'épuration est la difficulté de mesurer les paramètres nécessaires à une bonne régulation et relier à cela la nécessité de faire opérer ces usines par du personnel possédant une bonne formation tant du point de vue procédé que du point de vue instrumentation.

EXPANSION AND CONVERSION OF WINNIPEG'S EXISTING SECONDARY AERATION TREATMENT PLANT TO AN OXYGEN BASED COMPUTER-CONTROLLED PLANT

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INTRODUCTION

Sewage treatment exists in Winnipeg since the 1930s, at which time a primary plant had been constructed, together with two-stage anaerobic digestion. An initial extension to the primary plant took place in 1956 and secondary treatment consisting of an activated sludge air aeration plant became operational in 1965. This plant has a capacity of 250 mL/d, which has now been reached and further expansion is required. The City of Winnipeg, in a conceptual design report prepared by the Waterworks, Waste and Disposal Department, recommended that the plant be expanded to 330 mL/day. Their report further recommended that the existing air aeration plant be converted to an oxygen aeration system. This decision was based on favourable economics and also favourable experience with a smaller oxygen plant operating in the south part of Winnipeg for the last 10 years. Following the conceptual design, the City of Winnipeg hired the joint venture firms of Wardrop and MacLaren and subconsultants, EMA Inc., to further develop the conceptual design report into a functional design and, ultimately, into detail design.

My detailed discussion is divided into four major sections and computer technology is instrumental in the development of each. The message of "Computer-Controlled Plant" in the title of this paper is most meaningfull. This engineering assignment has relied on the computer for the design phase, and the resulting engineered design relies on the computer for proper operation. The construction staging was made possible because of the flexibility in computer control.

The four major sections are:

1. The conversion of the Secondary Treatment Process from air to oxygen;
2. The supply of oxygen;
3. The instrumentation and Control System;
4. The Plant Staging.

A. CONVERSION OF THE SECONDARY TREATMENT PROCESS FROM AIR TO OXYGEN

Figure 1 shows the existing facilities. One of the features of the existing plant is that the entire secondary treatment portion is covered. It was decided that the expanded plant would also be entirely enclosed for a number of reasons: protection from weather; ease of operation and control; and also, accessibility for maintenance. The City's conceptual design report recommeded conversion from an air-type plant to an oxygen-type plant. However, consultants were asked to further review this matter, particularly in view of new developments related to aeration, mainly the fine bubble dome-type of diffusers. The review took place considering two major factors: economic and process evaluation.

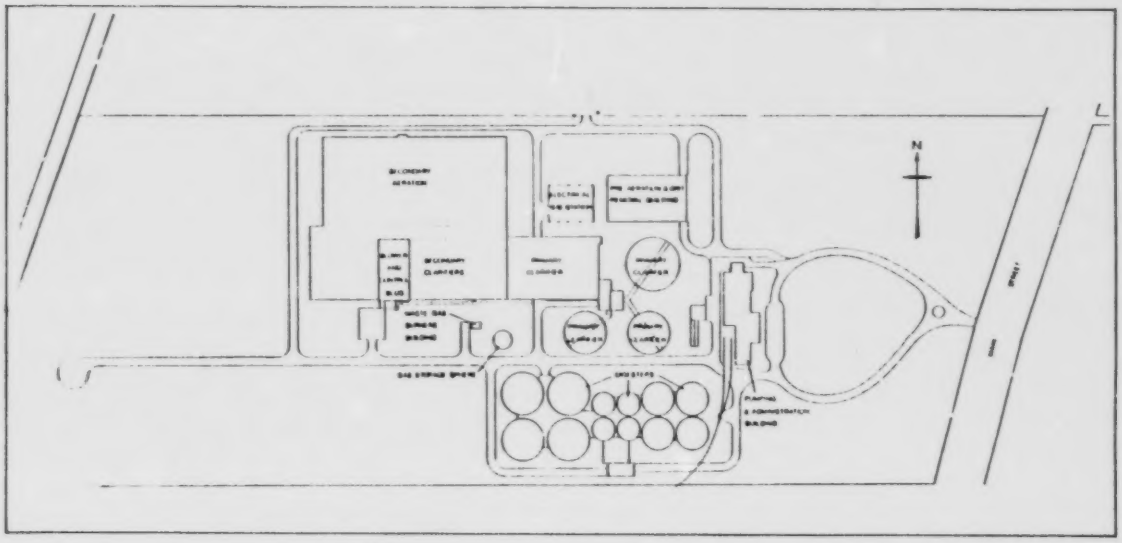

FIGURE 1 - EXISTING PLANT LAYOUT

Economic Considerations

The economic evaluation addressed four alternatives: two based on air and two based on oxygen.

Alternative 1 - The construction of new oxygen reactor tanks and conversion of the existing aeration tanks to final clarifiers.

Alternative 2 - The conversion of part of the existing aeration tanks to oxygen reactor tanks and converting the remaining aeration tanks to final clarifiers.This alternative also required additional final clarifiers to be built.

Alternative 3 - Construction of new, deep aeration tanks using fine bubble dome diffusers and conversion of the existing aeration tanks to final clarifiers.

Alternative 4 - Conversion of the existing aeration tanks with coarse bubble diffusers to fine bubble dome diffusers in conjunction with construction of additional final clarifiers. Additional new aeration tanks were also required under this scheme since conversion of the existing tanks did not provide sufficient aeration capacity.

Figure 2 shows the secondary treatment expansion, Alternatives 1 and 2, using the oxygen process.

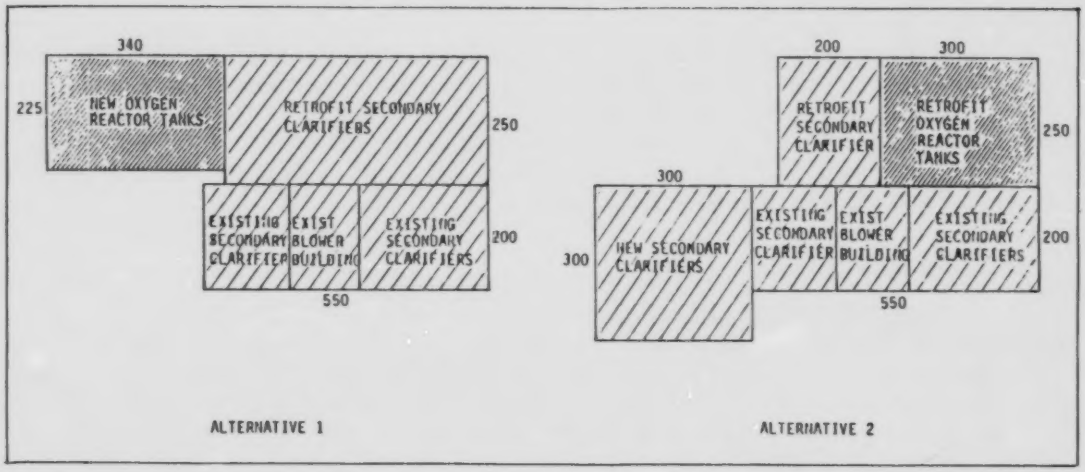

FIGURE 2 - SECONDARY TREATMENT EXPANSION ALTERNATIVES 1 AND 2 - OXYGEN PROCESS

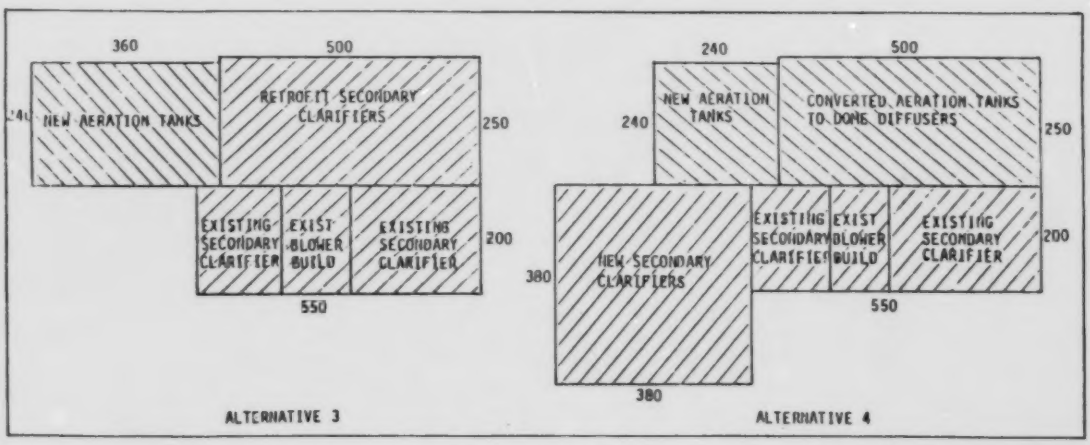

FIGURE 3 - SECONDARY TREATMENT EXPANSION ALTERNATIVES 2 AND 3 - AIR PROCESS

A review of Figs. 2 and 3 and comparing areas shows Alternative 1 and 3 involving new aeration/reactor tanks to be considerably more economical than Alternatives 2 and 4. One major reason is because of our design criteria required a superstructure over the final clarifiers while the aeration/oxygenation tanks can be covered only with an insulated roof slab.

The results of the economic analysis concluded that the capital and present value operating costs for the next 15 years clearly pointed out the advantages of the new tank alternatives. This is primarily because of the higher efficiency of the new deeper tanks. This was applicable to both air and oxygen alternatives. As a matter of comparison, the cost of the alternatives were:

Alternative 1 - Oxygen with new reactor tanks	$42.5M
Alternative 2 - Oxygen with converted reactor tanks	$58.4M
Alternative 3 - Air with new aeration tanks	$42.2M
Alternative 4 - Air with converted aeration tanks	$61.0M

The costs for oxygen alternatives included a City-owned and operated cryogenic oxygen plant. This item will be discussed later.

Process Evaluation

Since the cost of air and oxygen systems using the two new tank alternatives was very close, it was decided to evaluate the two process systems on other factors as well. Evaluation included plant visits and telephone surveys to various operating plants having experience with the two systems. The final step in the evaluation was the development of a matrix of 22 items to evaluate the other aspects such as process capabilities, and operating aspects. Three major classifications were used for evaluation.

1. Process and Operation Considerations
2. Maintenance Requirements
3. Other Considerations

A total of 250 points was agreed upon; then each member assigned a weight to each item considered. The results of this weighting was then averaged and tabulated as shown below.

ALTERNATIVE COMPARISON MATRIX
FOR SECONDARY PROCESS EXPANSION AT
THE NORTH END WATER POLLUTION CONTROL CENTRE

Comparison Parameters		Weighted Significance Final
Process and Operation Considerations	Group Total	134
Recovery from Shock Load		10
Process Reliability		41
Controlability		23
Instrumentation Complexity		14
Familiarity with Process		6
Operator Skill Requirements		11
Record of Experience		
Process		7
Operation		8
Operating Environment		7
Sludge Characteristics		7
Maintenance Requirements	Group Total	68
Maintenance Skills Required		15
Amount of Maintenance Necessary		24
System Downtime for Maintenance		20
Maintenance Environment		9
Other Considerations	Group Total	48
Odour Control		9
Safety		9
Process Warranty		5
Noise Potential		5
Energy Conservation		9
Site Use		3
Competitive Tendering		5
Use of Existing Equipment		3
TOTAL WEIGHTED GROUP RATING		250

You will note that of the three major classifications, the one on Process Considerations carried a weight of 134 out of a total of 250. This is so because plant performance and ease of operation was considered the main objective in selecting a treatment process.

The next step was for team members to assign points for each item on a preferential ranking system for oxygen and deep dome aeration. The total scores resulting from these steps are noted below for the three major classifications.

	Oxygen	Deep Dome
Process and Operation Considerations	3,819	3,203
Maintenance Requirements	1,647	1,889
Other Considerations	1,321	1,170
Total	6,787	6,262

You will note that the weighted result came out slightly in favour of the oxygen process. It is of interest to note that one category where the air system was favoured was in the maintenance requirements. The weighting team were concerned about the maintenance associated with operating a cryogenic plant. At that time, the intent was to have a City-owned and operated cryogenic plant; this was subsequently changed to privately-owned and operated plant as will be discussed later. If the oxygen alternative had initially been rated based on buying oxygen from a private utility, the oxygen alternative would undoubtedly have rated higher than it did.

B. THE SUPPLY OF OXYGEN

Background

As discussed earlier, it was concluded that a pure oxygen process was advantageous for the City. It was also concluded that the oxygen process offered process and operational advantages as well as reduced odour problems, less energy consumption and more effective use of the North End site.

Process

It was acknowledged that an oxygen system complete with oxygen generating facilities required more maintenance than an air system, and a higher level of skill would be required to operate and maintain the oxygen generating facility. Operation of a cryogenic plant requires sophisticated level of operation in a field not common in municipal installations.

Alternatives to City-owned and operated oxygen facilities were assessed. It was concluded that the City should purchase oxygen from a contractor rather than own and operate its own oxygen production facility. This decision was made after extensive investigations of the cost benefits, operation and maintenance requirements and overall dependability of the alternative oxygen supply schemes. Reaching this decision to have the contractor supply oxygen involved a lengthy process of discussion with suppliers and eventually, a tendering stage.

The City-owned option (make) was compared to the Contractor supply option (buy). This procedure is described more fully in the following paragraphs.

City-Owned Plant (make option)

A cryogenic oxygen generating plant with a capacity in the range of 86 tonnes/day was necessary to provide sufficient oxygen for the treatment process. The cost to construct a cryogenic plant was estimated at $10.5M in 1982 dollars. It was estimated the total present value of operating and maintaining the facility over the period from 1985 to 1999 would be approximately $6.1M (1982 dollars). The total estimated present value for the City-owned plant over a 15-year period was estimated at $16.65M.

Contractor-Owned Plant (buy option)

In the U.S. all or most municipal oxygen generating plants are owned by the municipalities. This is primarily due to the joint funding system where a large percent of the capital cost is funded by government. However, the majority of industrial oxygen users do not own the cryogenic facilities. They are owned and operated by an oxygen supply company which sells oxygen to the user or users by pipeline at a prescribed rate structure. Private users, of course, would have to fund the cost of owning and operating the oxygen generating facility entirely (as would the City of Winnipeg). It was determined, through discussions with the private sector that potential economic advantages existed by a privately-operated plant which could supply other markets or be dedicated to the North End. There was a strong interest displayed by the private sector to construct and operate a plant to satisfy the proposed Winnipeg oxygen requirements.

Comparison of Make and Buy Options

A comparison of the advantages and disadvantages associated with a contract oxygen supply versus a City-owned and operated plant was made. The results follow.

CITY-OWNED PLANT	CONTRACTOR-OWNED PLANT
Economic Considerations	
Requires considerable capital outlay.	Oxygen used is metered and purchased at a prescribed rate structure under the operating budget. No initial capital outlay is required.

CITY-OWNED PLANT	CONTRACTOR-OWNED PLANT (cont'd)
Economic Considerations (cont'd)	
No diversification. Results in higher costs to operate and maintain plant.	Possible economies associated with obtaining oxygen from an integrated merchant facility which supplies the marketplace as well as the City. Oxygen supply to the City from a "captive" plant (providing oxygen solely for the City) also offers economies in that the supplier has access to an integrated operation, maintenance and distribution network.
Economic Considerations	
Construction of plant requires design, tendering, construction by contract, with supervision by Engineer.	Could likely construct the oxygen plant more economically than the City. Most suppliers are vertically integrated companies who design, manufacture and construct the oxygen producing facilities, in addition to selling the oxygen to the marketplace.
Experience and Reliability of Supply	
No existing experience relating to the maintenance of cryogenic equipment.	Considerable experience in operating and maintaining equipment. Access to centralized parts inventory and computerized preventive maintenance programs.
No established track record relating to on-stream time.	97% on-stream time industry wide. Diversity of personnel to provide backup services for maintenance.
During down-time due to plant breakdown must negotiate purchase of oxygen in marketplace.	Guarantees to provide liquid oxygen at trucked prices during down-time due to plant breakdown. Has an oxygen distribution network with facilities for delivery of product and inventory in place to provide backup oxygen in the event of plant down-time.

Plant Characteristics

Standby compressor.	No standby compressor.
Single distillation column and reversing heat exchanger.	Single distillation column and reversing heat exchanger.

On the basis of the foregoing, a decision was made to call tenders for oxygen supply for the NEWPCC Secondary Treatment Expansion.

Basis of Tendering

As the City of Winnipeg owned a parcel of land adjoining the plant property, it was decided to make this site available for prospective oxygen vendors. The location of the City-owned plot of land relative to the oxygen reactor tanks is shown in Fig. 4 This figure also indicates the location of the successful tenderer's plant on the sewage plant property. (The successful tenderer submitted an alternate bid for a plant designated solely to provide oxygen for the North End Water Pollution Control Centre to be located on plant property.)

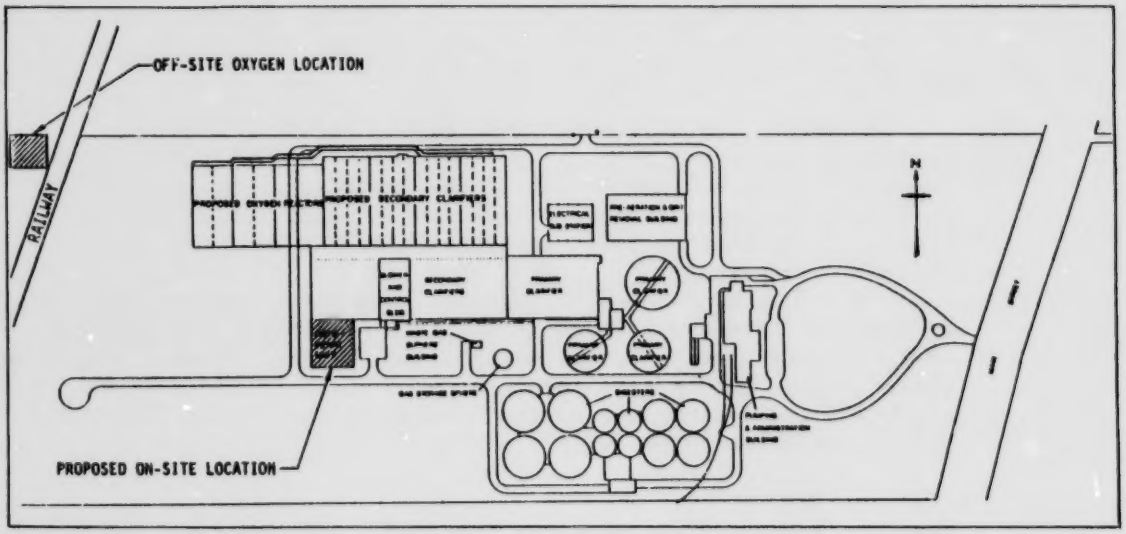

FIGURE 4 - LAYOUT FOR RENOVATED PLANT

Tenders for oxygen were called for at two supply pressures. This was required because two types of processes were considered to oxygenate sewage. One was the Unox/Oases process which uses low pressure oxygen supply at about 14 kPa. The second process was the Marox process which requires a higher oxygen supply pressure of 172 kPa. The oxygen

dissolution equipment was tendered under a separate contract which closed at the same time as the oxygen supply contract. The oxygen dissolution equipment tenders were evaluated in conjunction with the tenders for oxygen supply and took into consideration the varying costs of supplying oxygen at the higher pressure.

Due to the length of the purchase agreement, 15 years, and the many economic and physical factors which had a bearing on the cost, the tender evaluations were very complicated. Extensive discussions were held with oxygen suppliers throughout the development of the contract documents, as well as during the tendering phase. These discussions insured that tenderers had a full understanding of the terms and conditions of the contract and were able to conform to these conditions. The oxygen supply tenders were based on the following pricing structure:

1. A monthly base facility charge (fixed over the period of the contract). This charge provided for capital recovery and a return on the tenderer's original investment. The capital investment was taken to include:
   - the cost of the plant;
   - the cost of liquid oxygen storage facilities;
   - the cost of site services.
2. Indexed costs. These costs were indexed to various Statistics Canada indicators or other indices and included the following:
   - cost of energy;
   - cost of labour;
   - cost of maintenance.

Tenderers quoted the minimum take or pay oxygen charge, regardless of consumption. The minimum take or pay quantity was to account for the minimum output of the specific plant based on the maximum turndown conditions. The cost of oxygen requirements in excess of the minimum take or pay quantity was to be based on variable costs alone.

The tenderers were also required to quote a maximum gas consumption rate above which it would be necessary to supplement oxygen supply from a storage using liquid oxygen. Liquid oxygen was tendered at a higher rate than gaseous oxygen. Bidders were provided with anticipated oxygen demands so they could size their facilities to serve the City's needs.

Evaluation of Tenders

Four oxygen supply tenders were received. All tenderers possessed the necessary resources to satisfy the terms of the contract. In evaluating the oxygen supply tenders, the estimated present value of the cost to purchase oxygen from the various tenderers at the quoted prices

over a period of 15 years was compared to the present value cost for the City to own and operate a cryogenic oxygen plant over the same 15-year period. The tender evaluation also considered the effect of four factors upon the cost. These were: supply pressure; economic conditions; sewage flow variations; and, variable oxygen consumption rates.

The total cost of purchasing oxygen over 15 years is largely dependent upon projected flows and oxygen demands. As it is difficult to predict flows and demands accurately over a long period, a sensitivity analysis was undertaken to determine the effect of varying these factors on the total present value of each tender using the oxygen pricing formulas. As it was necessary to project oxygen consumption on an instantaneous basis for each day for 15 years through integration of the diurnal flow variation patterns and relative to the minimum take or pay quantities and the maximum gas flow rates tendered, a computer program was developed. The total cost in the year of expenditure for the oxygen utilized was calculated and converted to present value for each tendered case. These total present values were representative of the cost of oxygen over 15 years to the City of Winnipeg. Evaluation of the oxygen tenders with due consideration to variations of economic conditions, flow and oxygen consumption required examination of 64 cases for each tenderer, for a total of 256 cases for the base bid analysis. This evaluation was done by computer to save time and ensure accuracy. A sample of the computer printout resulting from the tender evaluation is shown in Fig. 5.

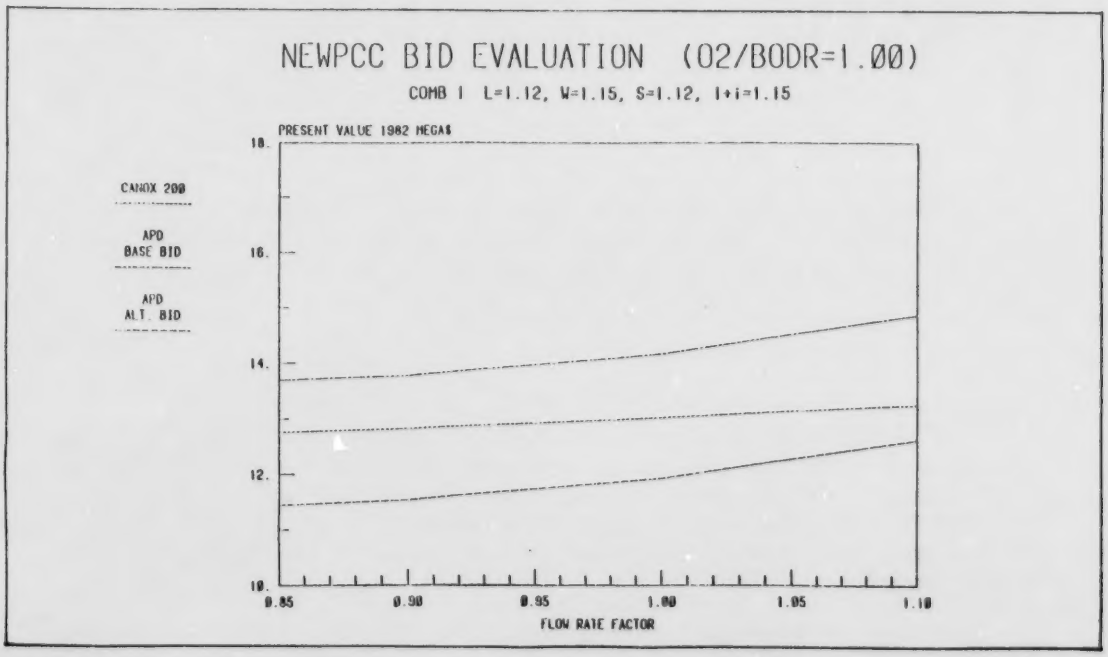

FIGURE 5 - COMPUTER PRINTOUT FOR OXYGEN SUPPLY TENDER EVALUATION

The comparison of the estimated present value, the cost to own and operate a City-owned plant (make option) was estimated to cost $16.6M. Buy option for the three lowest bids varied from $11.9M to $14.2M. One of the bids was submitted as an alternate by one of the tenderers. This bid was based on providing a dedicated plant on the site of the North End Sewage Plant rather than building the plant across the tracks on the original plot of land previously indicated. This alternative was eventually chosen which ends up being approximately $4.7M less costly on the present value basis than the City owning and operating their own facility.

Contract oxygen supply offers several economies as compared to the City-owned plant. Major cost savings are due to differences in the design of the facility itself and the reduction in quantity of liquid storage required in reserve on site to provide for the eventuality of the plant breakdown. All of the tenderers for oxygen supply were from large multi-national corporations with access to large distribution networks for liquid oxygen. All of these suppliers own tanker trucks and rail cars that could quickly transport oxygen to the site should a breakdown occur. This enabled the suppliers to provide only nominal standby equipment and still ensure continuous supply of oxygen through the distribution network.

In conclusion, it was found that the pipeline oxygen supply from a contractor-owned plant offered the City of Winnipeg a number of important advantages:

- substantial economies, approximately $4.7M over 15 years;
- continuity of oxygen supply, even during plant breakdown;
- simplified plant operation at reduced maintenance responsibilities.

The fact that the contractor supplied oxygen proved to be the most economical course of action for the City, further reinforced decision to utilize pure oxygen process. The major concerns relating to the oxygen process were related to the operating and maintenance aspects of a cryogenic oxygen plant.

The status now is that the plant has just been placed on stream, using the oxygen system. The cryogenic oxygen facility is completed and initial tests are currently being carried out on the actual process operation.

C. INSTRUMENTATION AND CONTROL SYSTEM

Introduction

In the City of Winnipeg's Conceptual Design Report, the City had recommended that the secondary treatment plant be fully automated using a Distributed Computer Based Process Control System. It was decided that as part of the Secondary Expansion, an overall Plant Control System would be designed now, making provision for addition in the future of other plant components.

Figure 6 shows the configuration of the proposed system.

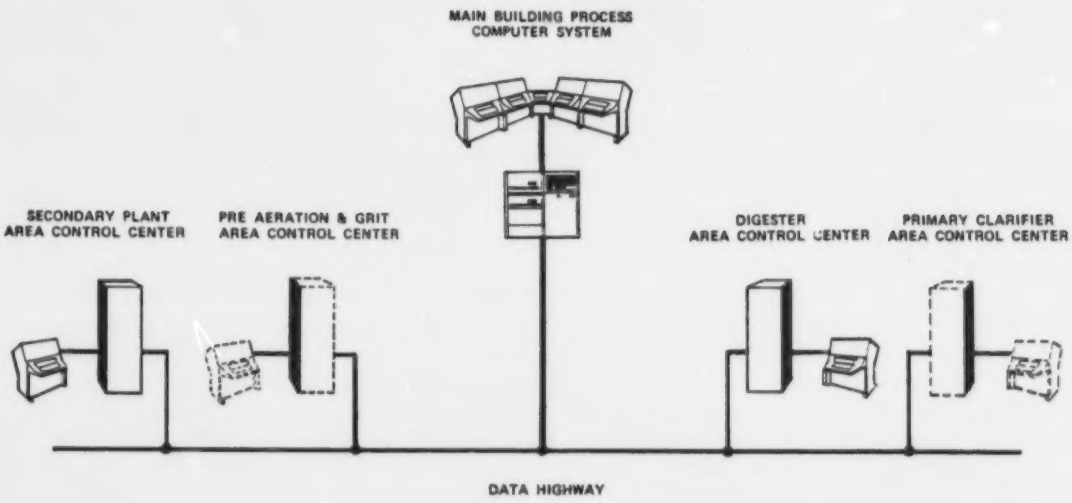

FIGURE 6 - CONFIGURATION OF THE PROPOSED INSTRUMENTATION AND CONTROL SYSTEM

The total system will initially include the main building computer system, the secondary plant local area control and the data highway sized for the complete system. The digester area control centre will be completed shortly thereafter.

Distributed Computer Based Control System

A definition of terms may be in order:

Distributed:	Not all in one place, divided, spread out.
Computer Based:	Using the technology of reconfigurable solid state or micro-electronic hardware.
Control:	The act of regulation by design or plan.
System:	An assembly which enables function as a unit.

Why Micro-Electronic Controls?

Treatment Plant Systems can be very complex and the secondary treatment plant portion in which we were specifically involved was particularly so. In selecting the control system, we established a requirement for accuracy, multiplicity and flexibility in the control system, and in our view, micro-electronics could provide the most suitable alternative.

Aptly designed micro-electronics can provide the plant operations with numerical control. Automatic control which formerly could only do the same routine over and over again, when displaced by numerical control, now enables the same routine to be modified on operator or machine command.

Good Control

The design team paid special attention to the factors of reliability and expandability while being very conscious of good operator interface and state of the art concepts.

Reliability

The distributed Computer Based Control System is the most reliable alternative because micro-electronics are inherently reliable.

Expandability

One of the significant aspects of control system philosophy is that a good system must be expandable.

The system architecture must readily accept increased tasks in an initially incorporated control scheme as well as permitting the orderly addition of more control schemes.

Expandability is the game because logic blocks and data base addresses can be added or deleted without any hardware changes.

Operator Interface

Operator interface is the main tool for proper execution. To execute his duties, the plant operator must be able to:

- See the process;
- Understand analog and digital signals;
- Interact with the control.

The computer based control system selected uses the CRT screen with color graphics to transmit all visual process information and a terminal keyboard for all operator interaction requirements. The old phrase "a picture is worth a thousand words" explains best why this selection was made.

State of the Art

Of all the design tasks involved with "controls", keeping up with the "state of the art" is the most fun, but also the most challenging. There is always something new and exciting to look at, but to select a system which will be in step with the "state of the art" in three to ten years is a serious challenge.

The first issue one has to accept is that hardware which is current now will be non-current or obsolete a number of years later. The key message here is the term non-current versus obsolete. Obsolescence of any hardware is not acceptable.

To select hardware components which are current today and only non-current later is one objective of the designer. A valuable step in the system selection process is that of locating suppliers (manufacturers) who are concerned that their equipment maintains the non-current status years later.

It was concluded during the design phase that the philosophy of the major instrumentation companies is to produce systems which are upward compatible and can be updated in a sequential fashion to remain current.

Control System Capabilities

The final system selected is not an "off the shelf" package but rather a tailor made assembly of standard components selected to meet very specific needs. The control capabilities identified for the NEWPCC in Winnipeg include:

1. Operation:
   - Decentralized sequence and regulatory control for area devices and subsystems (with centralized supervision).
   - Centralized monitoring for all plant processes and centralized reporting and logging of alarms and events.
   - Real-time display of process information in both alphanumeric and graphic form.

2. Management:
   - Advanced control capabilities for process control optimization.
   - Calculation of mass and flow balances, chemical and sludge inventories, loading rates, energy usage, etc.
   - Generation of real-time and historical trends and formal reports, of process and laboratory data.

3. Maintenance:
   - Scheduling of preventative maintenance functions based on equipment run time and/or calendar time.
   - Centralized monitoring and sequencial recording of process or equipment malfunction.
   - Generation of cost data for chemicals, labour and materials.

Two-Stage Tendering

In view of the complexity of this system, the changing state of the art, and to assure ourselves and the client that the hardware proposed is "state of the art" and likely, at worst, to become non-current rather than obsolete, the tendering process was a two-step procedure.

The first step consisted of a public tender requesting proponents to submit prequalifications in compliance with detailed plans and specifications. The system responsibility includes the supply of hardware and also the application software, the commissioning of system and operation availability demonstration.

The prequalification documents will be evaluated in detail by ourselves and the City. This will be followed by interviews to assure ourselves of proponent's capabilities. For the second round of tendering, only the qualified bidders approved in the first round will be requested to submit price proposals.

The current status is that the first round of tendering is closing this week.

Primary Elements

As a control system is only as good as the components used, the City of Winnipeg and ourselves have undertaken a series of tests of primary elements to determine which ones would be considered acceptable. Tests were carried out on:

- Dissolved Oxygen Analyzers
- Dissolved Oxygen Probes
- Suspended Solids Meters
- Sludge Blanket Detector
- Ultrasonic Flow Meters
- Variable Speed Drives

The major criteria for the test units are that they have the following characteristics:

- Wastewater does not foul the sensors
- Sample lines do not plug and are readily cleanable
- Solids in the sample do not interfere with transducers
- Simple calibration
- Infrequent calibration required

The specification documents only list those components that have proven satisfactory under the City tests, or similar tests done by others.

D. PLANT STAGING

The overall secondary plant expansion will be a 6-year project, from the beginning of the preliminary design phase to the end of construction. We, the consultants, obtained this assignment from the City of Winnipeg in early 1980 and the ultimate plant staging is intended to be completed by the end of 1986. The total plant expansion cost is expected to be in the order of $32M, that is, excluding the value of the cryogenic oxygen facility which is not directly a City capital expenditure. The overall program proceeded as follows. We, the consultants, obtained a conceptual design report from the City of Winnipeg which recommended an oxyen based computer operated plant. We further evaluated and confirmed the City's decision to proceed with an oxygen based plant. We also produced functional design drawings for the entire plant, together with a plant staging program. You will recognize that it is an important point that a sewage treatment plant cannot readily be interrupted while expansion takes place, so the construction program was staged to minimize shutdowns to brief tie-in periods.

Following approval by the City of the functional design and the overall estimate, detailed design was proceeded with on the entire project, the project was divided in 9 contracts, 5 equipment supply contracts and 4 installation contracts. The different contracts and the time frames for each are tabulated below:

CONTRACT SCHEDULES

Supply Contracts	Contract Award	*Contract Completion
Oxygen Supply	January, 1983	January, 1985
Oxygen Dissolution Equipment	January, 1983	January, 1985
Clarifier Mechanisms	June, 1984	*December, 1986
Pumping Equipment	*January, 1985	*December, 1986
Computer Control - Hardware and Software	*February, 1985	*March, 1987
Installation Contracts		
Administration Building	May, 1982	December, 1982
Reactor Tanks	January, 1983	January, 1985
Conversion of Aeration Tanks to Clarifiers	*February, 1985	*December, 1986
Computer Control System	*February, 1985	*March, 1987

*Estimated Dates

Current Status

The current status is that the oxygen supply plant has recently been commissioned by the Contractor. The oxygen process is now operational on part of the system and, in fact, the oxygen and air systems are operating in parallel. All oxygen reactor tanks and associated dissolution equipment are ready for operation, but only one has been placed in operation . Two of the four existing air-aeration tanks have been shut down, the other two are operating in conjunction with part of the existing final clarifiers. The remaining clarifiers are operating in conjunction with the oxygen reactors.

- The first stage of the two-stage tendering process for the computer control system is now underway.
- The pumping equipment is currently out for tender.

Summary

Phasing the entire project over a 6-year design and construction program has enabled an orderly conversion from the old to the new with very little interruption in treatment. It has also had the obvious advantage of spreading the $32M capital outlay over a period from 1982 to 1986.

OPERATIONAL AUDIT ASSISTS DESIGN OF PLANT MODIFICATIONS* FOR ENERGY SAVINGS AND IMPROVED CONTROL

by

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and

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Burlington, Ontario

ABSTRACT

A real time microcomputer-based operational audit was used to help select an activated sludge plant for a full-scale demonstration of the benefits achievable through the use of on-line instrumentation and automatic control. Information on aeration system performance and plant hydraulics was obtained to assess the suitability of a candidate plant. Off-line measurements of plant organic loading were also made. The dynamic data obtained indicated that significant savings could result with the implementation of continuous dissolved oxygen control. The candidate plant was therefore selected for the planned full-scale demonstration project.

INTRODUCTION AND BACKGROUND

More efficient operation of wastewater treatment systems can be achieved through the effective application of on-line instrumentation and automatic control. Process control, whether manual or automatic, is required because wastewater treatment systems are rarely, if ever, at steady state. This is true for the activated sludge process and particularly for its aeration component which directly influences energy expenditure and process performance.

In 1982, the Wastewater Technology Centre (WTC) commissioned a study to select a full-scale municipal wastewater treatment plant to demonstrate energy conservation and improved operation through automatic control. A consulting engineering firm screened four pre-selected plants ($<18 \times 10^3$ $m^3 \cdot d^{-1}$) in southern Ontario to assess their potential for being separated into two parallel aeration/clarification trains. Parallel trains were desired for comparison of conventional manual control to automatic control procedures. The results of the screening study were based largely on historical loading data and, to a lesser extent, on actual discrete measured data.

* Previously presented at and included in the proceedings of the 36th Western Canada Water and Sewage Conference, Regina, Saskatchewan, 19-20 September, 1984.

The outcome of the study was that the refit costs required to provide for parallel process trains were prohibitive at all of these plants. In addition, several concerns were raised about the approach used to evaluate the candidate plants. In particular, difficulties were identified with in situ aeration equipment performance evaluation, plant loading estimation, and flow split determination. It was decided that the screening procedure required re-assessment before proceeding with the evaluation of further candidate plants.

Concurrent to the initial screening exercise, research efforts on other projects yielded two plant monitoring tools that would be useful to investigate the performance of a plant under dynamic loading conditions. Both of these tools required the use of an Apple II microcomputer coupled to dissolved oxygen concentration and flow measurement instrumentation. The first tool was a software package developed in a joint program between the WTC and the University of Waterloo. This allowed for the in situ estimation of the overall oxygen mass transfer coefficient ($\alpha K_L a$) within an aeration tank. This parameter($\alpha K_L a$) is necessary to define oxygen transfer efficiency. The second tool, developed during an exchange program between the Danish Water Quality Institute (VKI) and the WTC, was an on-line instrument-based data acquisition software package for the Apple II microcomputer.

It was decided to use these tools to evaluate, in situ, the dynamics of the aeration system with respect to dissolved oxygen concentration, oxygen transfer efficiency, and hydraulic flow under actual operating conditions. This in-depth study was needed to estimate the potential energy savings achievable through automatic aeration control. Moreover, these data were essential to properly modify the process and the control system for a full-scale demonstration.

The Tillsonburg, Ontario WPCP was selected as the first candidate plant for evaluation using the newly-developed operational monitoring tools. Although the plant was smaller ($<9 \times 10^3$ $m^3 \cdot d^{-1}$) than originally desired, its hydraulic and air supply configurations appeared to be readily separable into two parallel process trains.

The Tillsonburg plant has two parallel trains (north and south) with two aeration tanks-in-series in each train. A common 0.03 m air header supplied air to both trains. The plant was fitted with twinned return and waste activated sludge pumps that could be isolated to provide one pump per train. One of the most attractive features of the plant was that two 45 kW and two 30 kW positive displacement blowers were available to deliver air to the common header. This represented apparent flexibility for control purposes if they could be split into equal pairs, each consisting of one 30 kW and one 45 kW blower. Each pair would then have been associated with either the north or south aeration train--the north being automatically controlled and the south manually operated for comparison.

MATERIALS AND METHODS

Microcomputer

An Apple II Plus microcomputer with memory expanded to 64K was used to perform on-line data acquisition for this study. Both Pascal and BASIC operating systems were required. The data acquisition program developed in Denmark was written in Pascal. This program used a "Mountain" real-time clock card for event scheduling. On-line sensors were interfaced to a "Mountain"

analog to digital (A/D) and digital to analog (D/A) converter card for signal acquisition. The aeration performance software developed at the WTC and University of Waterloo was written in BASIC and used an Interactive Microware Inc. "ADALAB" real-time front-end micro-processor interface with an integral real-time clock and A/D converter. The "Mountain" clock card and both interface cards were installed in the internal slots of the microcomputer and standard connectors were used to switch the sensor signals to the respective A/D interface card to facilitate running either the data acquisition or the aeration performance program.

Peripherals included a thermal printer plotter, monochrome monitor and two 5-1/4" floppy disk drives.

Dissolved Oxygen Measurement

A standard laboratory model Yellow Springs Instrument Company (YSI) dissolved oxygen (D.O.) measuring system was used to monitor the D.O. concentration in each of the four aeration tanks. Specifically, two Model 54 and two Model 54A transmitters were used. A Model 5750 YSI probe complete with a 15 m signal cable was connected to each transmitter. Each probe was weighted and supported by a cable running the length of the respective aeration tank. This allowed flexibility in terms of probe placement. The probes were placed at mid-depth, one third down the length of each first pass and two thirds down the length of each second pass.

Output signal conditioning, isolation and amplification were accomplished at each YSI transmitter using separate 4-20 mA output signal converter/isolation transmitters. Careful consideration of installation, grounding and power supply was required to ensure good overall performance. Temporary, 50 m, shielded twisted pair signal wires were installed from each isolation transmitter to the microcomputer. All YSI transmitters, the 117 V a.c. power supply, and signal isolation transmitters were located in a plywood housing bracketed to the handrail of one tank as shown in Figure 1.

FIGURE 1. DISSOLVED OXYGEN SIGNAL TRANSMISSION EQUIPMENT.

Continuous D.O. monitoring required continual recharging of the Ni-Cad rechargeable cells in each YSI transmitter. To accommodate this, each transmitter had a built-in charger designed for 117 V a.c. operation. Initially, the four probes were calibrated, bundled together and immersed in a plastic bucket to establish their consistency. Excellent agreement, within 0.1 $mg \cdot L^{-1}$, was observed. An attempt was made to establish that the consistency was maintained within one of the aeration tanks. When the bundled probes were immersed in an aeration tank, it was observed that the response of the Model 54 transmitters deviated significantly (>1.0 $mg \cdot L^{-1}$) from that of the Model 54A transmitters. This was thought to be due to a difference in ground potential between the aeration basin and the two models of transmitters. This was eliminated by jumpering the 117 V a.c. recharger within each Model 54A transmitter directly to the charging circuit within a Model 54 transmitter. This provided a common ground for both the Model 54A and Model 54 transmitters.

The probes were air calibrated using a standard BOD bottle as outlined in the YSI user manual at approximately weekly intervals. The calibration rarely needed adjustment of more than 0.3 $mg \cdot L^{-1}$. The membrane and electrolyte were changed when the signal became obviously erratic. This usually was sufficient to re-establish the measurement. On one occasion, moisture was found to have penetrated a splice in the probe cable and the symptoms were similar to those suggesting membrane replacement.

Level Measurement

The secondary clarifier effluent from the north train of the plant passed through a 0.023 m Parshall flume. A 0.061 m contracted rectangular weir was used to measure the south train effluent. Robertshaw Model 157 "Level Tel" capacitance probes were used for liquid level measurement in each of these devices. Liquid level signals (4-20 mA) were transmitted over temporary, 50 m, shielded twisted pair signal cables to the microcomputer where on-line calculation of actual flow rates was made and recorded. A third level detector was installed at the grit chamber discharge weir to monitor the operating characteristics of the upstream pumping station. This was necessary because it was expected that the plant hydraulic characteristics would dampen the input hydraulic disturbances prior to measurement at the effluent flume and weir. These data would be required if necessary, to provide for modification of the pump station hardware and control mechanism.

Off-line Sampling

Automatic samplers were used to collect discrete samples at 3-hour intervals on a 24-hour basis for both primary and secondary effluent of each process train. Table 1 lists the constituents measured.

TABLE 1. OFF-LINE ANALYSES FOR PRIMARY AND SECONDARY EFFLUENT

Unfiltered	Filtered
BOD_5	TKN
COD	NH_3/N
TKN	NO_2/N
SS	NO_3/N

Data Acquisition

The on-line instrument signals were sampled every 10 seconds and the data displayed in engineering units in tabular form on the monitor screen. At 6-minute intervals, the data were time stamped, and logged in integer format to a floppy diskette. The sample and logging intervals were user selectable. Software was written to allow the stored data to be plotted graphically on the monitor or printer/plotter. However, while the plotting program was executing, data acquisition ceased because of the single tasking nature of the computer.

Aeration Efficiency Determination

The aeration transfer efficiency was measured using a non-steady state reaeration technique. This was done at 30 kW and 75 kW blower configurations using the BASIC program in the microcomputer to collect D.O. data at 90-second intervals after a known step change in blower power. The details of this procedure have been described elsewhere (Howell *et al*, 1984).

RESULTS AND DISCUSSION

When the monitoring program was started up in August 1983, several unusual plant operating features were uncovered. From the hydraulic measurements, the north train of the plant was found to be receiving about 20% more flow than the south train. This difference is shown in Figure 2, which depicts the flow split during a 24-hour period. Moreover, it was shown that the flow to the plant was never at a steady value due to the upstream lift station control characteristics. Figure 3(a) shows that the grit tank level oscillated from zero to maximum with a 6-minute period due to the on/off pump station control being exercised. The dampening effect that the plant has on the given input hydraulic disturbances is evident from the corresponding effluent weir flows shown in Figure 3(b).

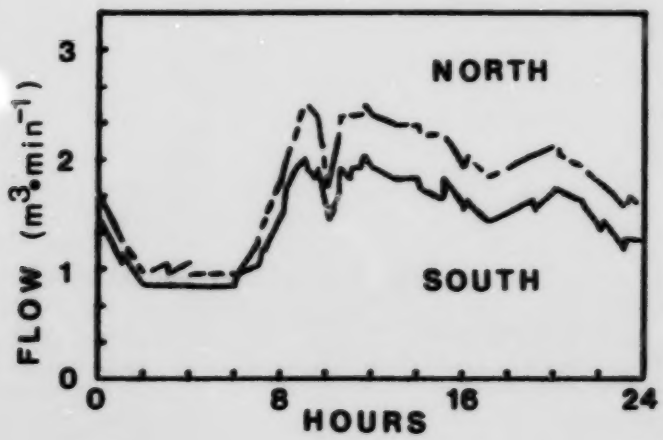


FIGURE 2. EFFLUENT HYDRAULIC FLOW MEASUREMENT.

These disturbances may be large enough to have a significant effect on the performance of the secondary clarifiers. Therefore, prior to the demonstration program, the flow split may have to be balanced and perhaps the pump station control approach will require modification in order to carry out the study. The dynamic data illustrated in Figures 2 and 3 will be used to determine how to utilize the future plant computer for pump control to minimize the hydraulic disturbances to the plant.

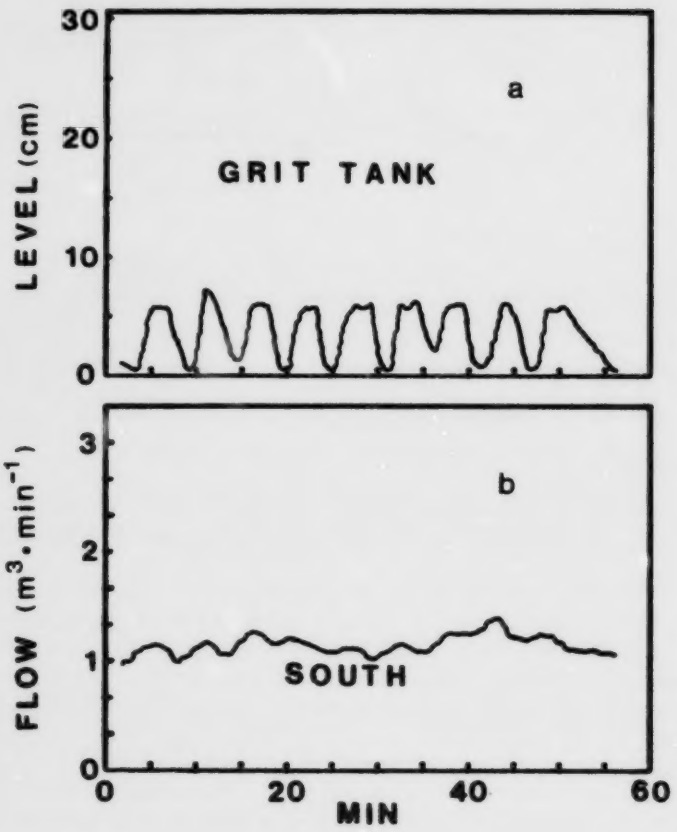


FIGURE 3. INFLUENT (A) AND EFFLUENT (B) HYDRAULIC DISTURBANCE CREATED BY ON/OFF PUMP STATION OPERATION.

The continuous on-line monitoring of the aeration tanks produced some interesting results in connection with D.O. concentration. With only 45 kW blower capacity in operation and approximately equal air distribution to the four tanks, the D.O. concentration was observed to be essentially zero most of the day (Figure 4). The two left hand graphs in Figure 4 represent the first pass of each train while the right hand graphs represent the second pass. It is apparent that the load to the two first passes exceeded their oxygenation capacity at the 45 kW blower output level. Since most of the BOD removal occurred in the front passes, the second passes achieved higher D.O.'s for

longer periods, but could not maintain a residual D.O. except at minimum loadings.

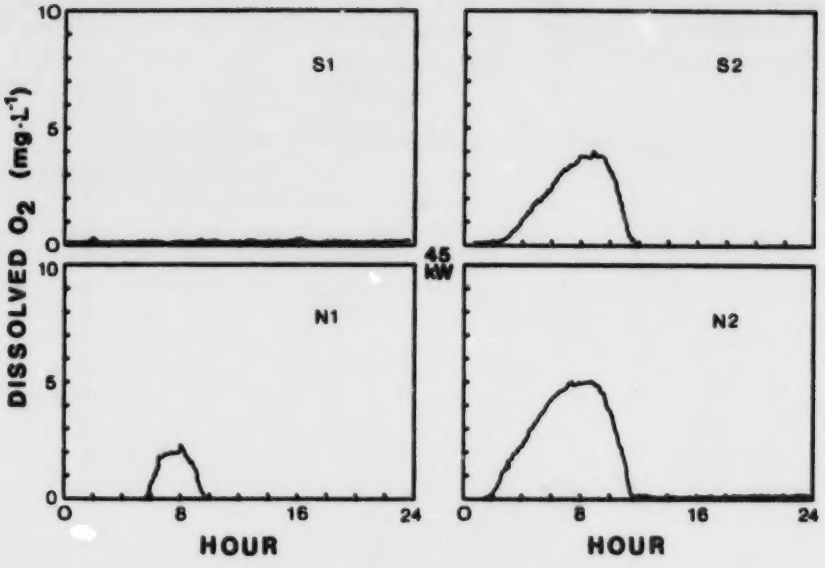


FIGURE 4. DISSOLVED OXYGEN CONCENTRATION AT 45 kW BLOWER CONFIGURATION AND ORIGINAL AIR DISTRIBUTION.

Throughout this time, the visual quality of the plant effluent was excellent. In retrospect, it is believed that nitrification was controlling the D.O. under these conditions. Plant operating records indicated partial nitrification was occurring, presumably to the extent allowed by the transfer of oxygen into the system, i.e., nitrification was oxygen-limited at peak organic loadings.

Positive D.O. measurements were required at all times to permit the necessary assessment data to be collected. Therefore, the blower output was increased to the 75 kW configuration by starting one of the 30 kW blowers. This immediately resulted in an increased D.O. in all of the tanks during the off-peak loading period. The D.O. concentration was higher for a longer period of time in the second passes than in the first passes. However there was zero D.O. in the tanks for significant periods of the day even under the 75 kW configuration (Figure 5). The 30 kW blower was then shut off after it became apparent that the increased blower capacity alone did not achieve the desired D.O. profile.

After the 75 kW trials, the airflow was redistributed since it was recognized that the second passes needed less air than the first passes. The air flow to the second passes was reduced to sustain mixing only, to force more air to the first passes. With 75 kW blower capacity in operation this achieved the results shown in Figure 6. Although highly dynamic variations in D.O. are apparent, zero D.O. concentrations did not occur. Nitrification was known to be complete under these conditions.

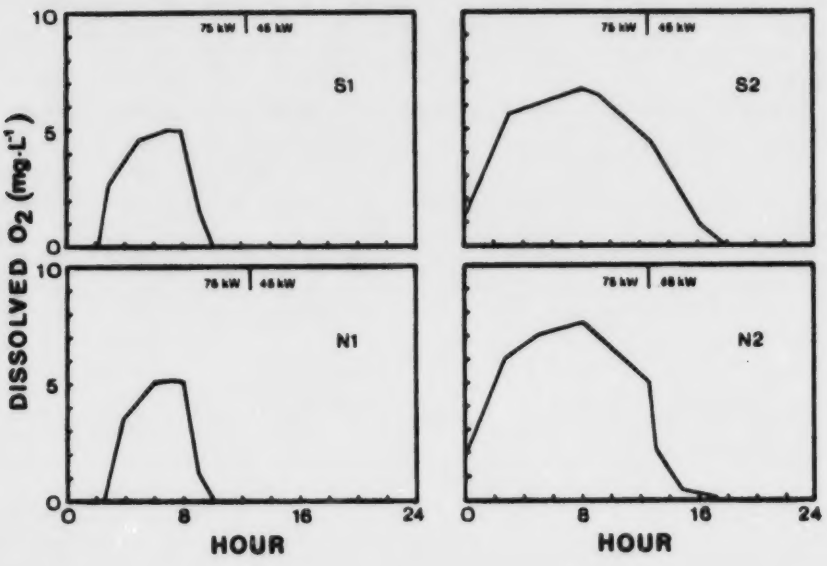


FIGURE 5. DISSOLVED OXYGEN CONCENTRATION AT 75 kW BLOWER CONFIGURATION AND ORIGINAL AIR DISTRIBUTION.

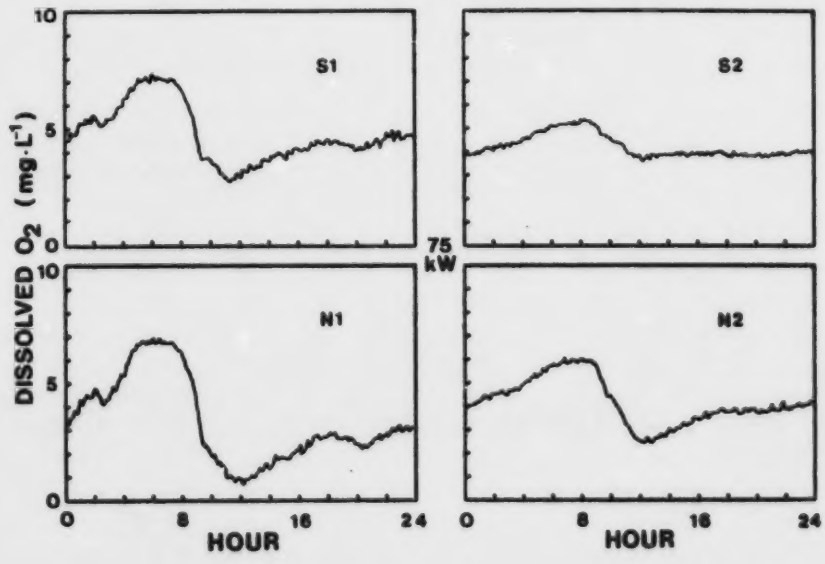


FIGURE 6. DISSOLVED OXYGEN CONCENTRATION AT 75 kW BLOWER CONFIGURATION AND REDISTRIBUTED AIR FLOW.

A measure of the overall oxygen mass transfer coefficient, $\alpha K_L a$, was made at total blower power configurations of 30 kW and 75 kW. The air distribution was the same as that which achieved the results shown in Figure 6. The results from the $\alpha K_L a$ determinations are given in Table 2.

TABLE 2. OVERALL OXYGEN MASS TRANSFER COEFFICIENT, $\alpha K_L a$

Total Blower Configuration (kW)	Aeration Tank			
	South 1 (h^{-1})	North 1 (h^{-1})	South 2 (h^{-1})	North 2 (h^{-1})
30	1.15	1.38	1.48	0.44
30	1.39	1.17	-0.34	0.64
75	3.18	2.81	1.68	1.25
75	4.22	3.26	1.48	0.89
75	4.11	4.54	1.40	2.24

For each tank, it was assumed that the relationship between $\alpha K_L a$ and the total blower power configuration was linear and passed through the origin. A least squares analysis was used to define this relationship. For example, in tank South 1:

$$\alpha K_L a = 0.05030 \text{ (blower power)} + 0.0 \qquad (1)$$

The family of curves generated for the four aeration tanks is shown in Figure 7. The ratio of the $\alpha K_L a$ estimated for each tank to the sum of the $\alpha K_L a$'s for all four tanks at a given blower configuration was used to obtain an estimate of the individual air flow being supplied to each tank under process conditions. For example, at the 75 kW blower configuration, the proportion of total airflow supplied to aeration tank South 1 was estimated to be:

$$\% \text{ airflow to S1} = \frac{\alpha K_L a \text{ (South 1)} \times 100}{\sum \alpha K_L a\text{'s (four tanks)}} = \frac{3.77\ h^{-1} \times 100}{10.23\ h^{-1}} = 36.9\% \qquad (2)$$

Similarly, the airflows to tanks N_1, S_2, and N_2 were estimated to be 34.2%, 14.8% and 14.2% respectively. The total airflow for a given blower configuration was determined from the factory blower performance curves.

From D.O. data similar to those shown in Figure 6 and knowledge of the relationship between the overall oxygen mass transfer coefficients and blower power (Figure 7), the power consumption under simulated D.O. control conditions was estimated. This required a mass balance on the mixed liquor D.O. expressed as:

$$dC/dt = \alpha K_L a\ (C_s - C) - R \qquad (3)$$

where C = dissolved oxygen (D.O.) concentration ($mg \cdot L^{-1}$)
C_s = saturation value of D.O. at process conditions ($mg \cdot L^{-1}$)
$\alpha K_L a$ = overall oxygen mass transfer coefficient (h^{-1})
R = oxygen utilization rate (OUR) or respiration ($mg \cdot L^{-1} \cdot h^{-1}$)

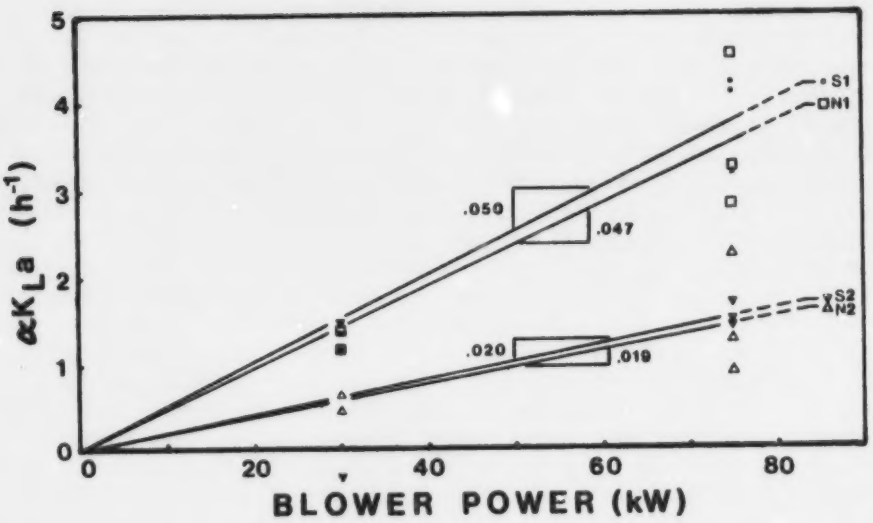


FIGURE 7. RELATIONSHIP OF OVERALL OXYGEN MASS TRANSFER COEFFICIENT ($\alpha K_L a$) AND BLOWER POWER CONFIGURATION.

The respiration rate, R, could be calculated from Equation 3 since all the other parameters were known for the conditions under which the on-line D.O. data were collected. However, it was necessary to recognize that the D.O. concentration was never at steady state and that the derivative term in Equation 3 was not negligible. Therefore, Equation 3 was discretized as follows:

$$(C_t - C_{t-1})/\Delta t = \alpha K_L a_{t-1} \quad (C_s - C_{t-1}) - R_{t-1} \tag{4}$$

where Δt = sample time interval
t = current sample time
$t-1$ = previous sample time

The respiration rate was calculated by rearranging Equation 4 to:

$$R_{t-1} = \alpha K_L a_{t-1} \quad (C_s - C_{t-1}) - (C_t - C_{t-1})/\Delta t \tag{5}$$

Once R was estimated for the actual operating conditions, it was possible to simulate the total blower power which would be required to maintain the dissolved oxygen concentration at a setpoint level. With setpoint control, the D.O. concentration is constant with time and Equation 5 can be rearranged to give:

$$\alpha K_L a_{t-1} = \frac{R_{t-1}}{(C_s - C_{t-1})} \tag{6}$$

Equation 6 was used to calculate the $\alpha K_L a$ values required to maintain the setpoint D.O. concentration under variable loading conditions for each

tank. From the calculated $\alpha K_L a$ values, and the relationship between $\alpha K_L a$ and total blower power for each tank as shown in Figure 7, estimates of the total blower power required to maintain the D.O. setpoint in the respective tank were made. This had to be done separately for each tank since only the total blower power required to deliver air to all tanks through the common header could be measured.

The portion of blower power attributable to air delivery for each individual tank could not be measured directly. Therefore, the power required for setpoint D.O. control in each tank was estimated as if only that tank, with its variable loading and fixed air distribution valve position, were controlling the total blower configuration. Figure 8 presents simultaneously four different examples of the power consumed by the whole system for the four cases in which the D.O. level in the respective tank is controlled at 1.0 $mg \cdot L^{-1}$. Only the tank used to control the total blower output would maintain its dissolved oxygen concentration at the setpoint value. The D.O.'s measured in the other tanks would vary from their respective setpoint values if the total blower power requirement determined by the controlling tank was different from that calculated to be required for setpoint maintenance in the other tanks. For example, to maintain a D.O. of 1.0 $mg \cdot L^{-1}$ in tank North 1, the total system power draw was calculated to be 1170 kWh by integrating the area under the power curve.

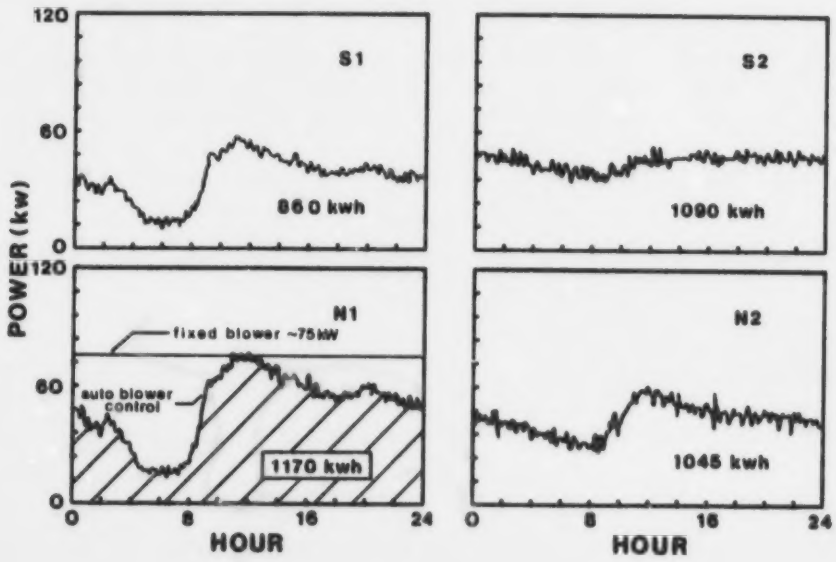


FIGURE 8. TOTAL SYSTEM BLOWER POWER CONFIGURATION FOR SIMULATED 1.0 $mg \cdot L^{-1}$ SETPOINT D.O. CONTROL (SEPARATE CASE FOR EACH TANK).

Without control (i.e., constant blower output), the power requirement to maintain at least 1.0 $mg \cdot L^{-1}$ D.O. in tank North 1 would have been 75 kW - resulting in a 24 hr energy consumption of 1800 kWh. The reduction in aeration energy through automatic control (1800-1170 kWh) represents approximately 30% savings over that consumed by fixed blower operation. This is

likely a conservative estimate because air distribution control would also minimize the air header pressure while controlling the D.O. in all tanks, thus further reducing the energy consumption for the total system.

This analysis indicated that there was a strong possibility of demonstrating substantial energy savings using aeration control at the Tillsonburg plant. Therefore, the plant was selected for the purpose of demonstrating at full-scale the potential for improved operation and energy savings through the use of on-line instrumentation and control. Although the details of the refit design are beyond the scope of this paper, the computer-based, on-line audit provided much of the data upon which the design was based. Other factors such as improved biological solids inventory control through automation will also be investigated to determine the benefit in relation to improved plant operation and energy savings.

SUMMARY AND CONCLUSIONS

The Wastewater Technology Centre demonstrated the use of a portable, microcomputer-based, real-time monitoring system for assessing the operational capability of an existing full-scale activated sludge plant.

The real-time dynamic operating data obtained through the use of the audit equipment were valuable in determining the existing capacity of the aeration equipment and the present operating conditions at the plant. This type of continuous data is new to plant operators and those considering plant refits or expansions. It can provide the necessary dynamic information for correct specification of process hardware such as blowers, control valves and diffusers, to provide for operational flexibility under dynamic operating conditions.

A study of the plant aeration energy requirements demonstrated that to maintain a minimum D.O. concentration of 1.0 $mg \cdot L^{-1}$, an estimated 30% aeration energy could be saved if the plant was refitted for aeration control, as compared to fixed blower operation.

The on-line operational audit is a valuable tool which can provide much more process information, with less effort, than can be obtained through manual or discrete sampling.

ACKNOWLEDGEMENTS

This work was financially supported by Environment Canada and the Office of Energy Research and Development, Energy Mines and Resources Canada. Special acknowledgement is extended to the Ontario Ministry of Environment and the Tillsonburg WPCP staff in particular for use of their facilities. Wastewater Technology Centre personnel who contributed significantly to this work included: Eric Luxon, Jim Matthews, Glen Horn, Mark Yendt, and Ron Gillespie.

REFERENCES

1. Howell, J.A. et al., "On-line measurement of respiration and mass transfer rates in an activated sludge aeration tank", Journal of the Water Pollution Control Federation, 56, (4), 319-325, 1984.

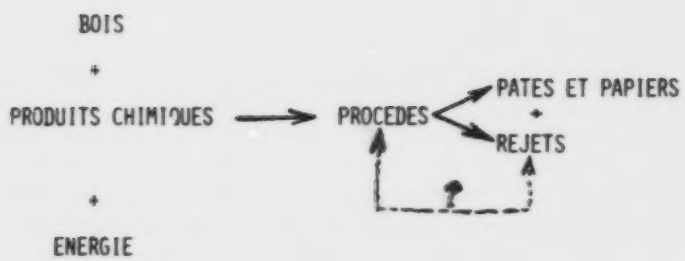
MÉTHODOLOGIE DE DÉPOLLUTION DANS LES USINES DE PÂTES ET PAPIERS DU QUÉBEC

H.C. Lavallée
Ministère de l'Environnement du Québec

OBJECTIFS

- RESUMER LE TRAVAIL ACCOMPLI
- PRESENTER LES ASPECTS TECHNOLOGIQUES PARTICULIERS POUR LE TRAITEMENT DE LA POLLUTION RESTANTE

USINES DE PATES ET PAPIERS



USINES DE PATES ET PAPIERS

PROCEDES $\xrightarrow{?}$ REJETS

BOIS	ETAPES	REJETS
PREPARATION	- LAVAGE	FIBRES, MATIERES DISSOUTES
	- ECORCAGE	FIBRES, MATIERES DISSOUTES, SUBSTANCES TOXIQUES
PATE	- DEFIBRAGE*	MATIERES DISSOUTES, SUBSTANCES TOXIQUES, FIBRES, pH
	- BLANCHIMENT	MATIERES DISSOUTES, SUBSTANCES TOXIQUES, FIBRES, pH
	- EPURATION	FIBRES
PAPIER	- FABRICATION DU PAPIER (M.P.)	FIBRES, MATIERES DISSOUTES
	- TRANSFORMATION	FIBRES

* ELEMENT DETERMINANT POUR LA CLASSIFICATION D'UNE USINE

USINES DE PATES ET PAPIERS

PROCEDES (DEFIBRAGE)

CATEGORIES		QUEBEC
PROCESSUS CHIMIQUES:	PATE, PATES ET PAPIERS, PATE ET CARTONNAGE:	28 USINES
PROCESSUS MECANIQUES:	PAPIERS FINS, PAPIERS TISSUS, AUTRES PRODUITS P & P:	31 USINES

CARACTERISATION GLOBALE

UTILISATION D'EAU: 2.5×10^6 m^3/D (550×10^6 MGID) - 77% CATEGORIE CHIMIQUE

- IDEM QUANTITE D'EAUX DOMESTIQUES DU QUEBEC -

NATURE DE LA POLLUTION (REJETS)

POLLUANTS CONVENTIONNELS (REGLEMENTES)

- M.E.S. (MATIERES EN SUSPENSION. E.G. FIBRES)
- DBO_5 (MATIERES DISSOUTES)
- pH

POLLUANTS NON CONVENTIONNELS ET TOXIQUES

- BPC
- ACIDES RESINIQUES ET GRAS
- DERIVES DU BLANCHIMENT
- COMPOSES PHENOLIQUES (TCP. PCP. ETC.)
- CHLOROFORME
- COULEUR
- ECUME

REGLEMENTATION DES PATES ET PAPIERS

- EXCLUSIVE AU QUEBEC (A.C. 2349-79)
- EN VIGUEUR DEPUIS SEPTEMBRE 1979
- NORMES SUR LES EFFLUENTS. EMISSIONS ET DECHETS SOLIDES
- ECHEANCIERS A RESPECTER
- ADAPTEE A DE VIEILLES USINES. CONFINEES ET NOMBREUSES (59)
- GRANDE DIVERSITE DE PROCEDES DE FABRICATION
- APPORT ECONOMIQUE IMPORTANT

NORMES ET ECHEANCIERS

- NORMES: KG/T (% PRODUCTION) - CHARGE MOYENNE MENSUELLE
  - CHARGE MAXIMALE QUOTIDIENNE

- ECHEANCIERS

	USINES EXISTANTES	USINES NOUVELLES
POLLUANTS CONVENTIONNELS (RÉGLEMENTÉS)	M.E.S. - FIN 1983	IMMÉDIAT
	DBO_5 - 1ÈRE ÉTAPE - FIN 1986 (PROPOSÉE)	IMMÉDIAT
	DBO_5 - 2E ÉTAPE - *	IMMEDIAT **
POLLUANTS NON CONVENTIONNELS ET TOXIQUES (NON RÉGLEMENTÉS)	PROTOCOLE D'ENTENTE *	IMMEDIAT **

* SELON LE PROGRAMME D'ASSAINISSEMENT DES EAUX (PAE) - BESOINS DU MILIEU AQUATIQUE

** SELON UNE ÉTUDE ENVIRONNEMENTALE ET PAE

SUBDIVISIONS DES CATEGORIES SELON LE REGLEMENT

CATEGORIES	OPERATIONS	ELEMENTS DE TRANSFORMATION	NOMBRE
I	PRÉPARATION DU BOIS	LAVAGE ET ECORCAGE	2
II	PRÉPARATION DES PÂTES	PÂTES CHIMIQUES KRAFT, AU BISULFITE (3), MI-CHIMIQUE ET A DISSOUDRE PÂTES MÉCANIQUES ET CHIMICO-MÉCANIQUE BLANCHIMENT DES PÂTES PÂTE EN FEUILLE	13
III	FABRICATION DU PAPIER	INTÉGRÉE ET NON-INTÉGRÉE UN OU PLUSIEURS PRODUITS AVEC OU SANS RECYCLAGE DES EAUX BLANCHES AVEC OU SANS ADDITIF CARTON ET AUTRES	17

NOTES: 1) CHAQUE ÉLÉMENT DE TRANSFORMATION PORTE UNE NORME SPÉCIFIQUE
2) LA NORME GLOBALE D'UNE USINE EST CONSTITUÉE DE LA SOMME DE CHACUNE DES NORMES VISANT LES ÉLÉMENTS DE TRANSFORMATION OPÉRANT A CETTE USINE

SIGNIFICATION TECHNOLOGIQUE DES NORMES

M.E.S. (TOUTES LES USINES)

- TRAITEMENT PRIMAIRE PAR SEDIMENTATION
  - AVEC OU SANS ADDITIF
  - EFFICACITE MOYENNE DE 80%
- RECUPERATION DES BOUES
  - INCINERATION (47%)
  - DISPOSITION (51%)
  - RECYCLAGE (2%)

SIGNIFICATION TECHNOLOGIQUE DES NORMES

DBO_5 - 1ère ETAPE

- PROCESSUS CHIMIQUES	- RECUPERATION DES LESSIVES DE CUISSON - CHANGEMENT DE PROCEDE
- PROCESSUS MECANIQUES	- REJETS BRUTS AUTORISES

DBO_5 - 2e ETAPE

- TOUTES LES USINES	- TRAITEMENT BIOLOGIQUE

pH

- TOUTES LES USINES	- NEUTRALISATION

M.E.S.

OPTIONS TECHNOLOGIQUES

1- TRAITEMENT PRIMAIRE GLOBAL (P.G.)

AVANTAGES	DESAVANTAGES
CONTROLE EN UN POINT	COUTEUX
CAPACITE TAMPON	ESPACE IMPORTANT
SECURITAIRE	SYSTEME D'URGENCE REQUIS
	"EFFET DE POUBELLE"
	GRAND VOLUME DE BOUES

M.E.S.

OPTIONS TECHNOLOGIES

2- TRAITEMENT PRIMAIRE PARTIEL (P.P.)

AVANTAGES	DESAVANTAGES
SUR EFFLUENTS CONTAMINES	SEGREGATION INTERNE COUTEUSE
MOINDRE COUT GLOBAL	DOUBLE SYSTEME - INT. ET EXT.
OPERATIONS D'ECORCAGE	
CERTAIN RECYCLAGE DE FIBRES	

M.E.S.

OPTIONS TECHNOLOGIQUES

3- TRAITEMENT INTERNE GLOBAL (I.G.)

AVANTAGES	DESAVANTAGES
OPTIMISE LES PROCEDES	SURVEILLANCE CONSTANTE
REDUIT CONSOMMATION D'EAU	INSTRUMENTATION ELABOREE POUR PROCEDES SIMPLES
ECONOMISE ESPACE ET ENERGIE	ENTRAINEMENT ET FORMATION DU PERSONNEL
MEILLEURE RENTABILITE	RISQUE ELEVE PERTES ACCIDENTELLES
RECUPERATION A LA SOURCE	

M.E.S.

ANALYSE DU CHOIX DES OPTIONS

OPTIONS	AVANT 1980	APRES 1980
1- P.G.	14	11
2- P.P.	1	7
3- I.G.	2	14
TOTAUX:	17	32
GRAND TOTAL:	49	

M.E.S.

ANALYSE DU CHOIX DES OPTIONS

OPTIONS	AVANT 1980	APRES 1980
1- P.G.	14	11
2- P.P.	1	7
3- I.G.	2	14
TOTAUX:	17	32
GRAND TOTAL:	49	

CONCLUSION:

P.G. (TRADITIONNEL): 82% → 34%

P.P. + I.G. : 18% → 66%

M.E.S.

CONFORMITE AU REGLEMENT (OCTOBRE 1984) PAR OPTION

- CONFORMITE

OPTIONS	AVANT 1980		APRES 1980		TOTAL	
	C	NC	C	NC	C	NC
1- P.G.	12	2	8	3	20	5
2- P.P.	1	0	5	2	6	2
3- I.G.	2	0	3	11	5	11
TOTAUX:	15	2	16	16	31	18
	17		32		49	
- REJETS (T/D)	600		300		DIMINUTION 300 (50%)	
- COUTS (MM$) (APRÈS 1980)	DÉPENSÉ: 130 EN COURS: 40		170*			

* SUBVENTIONNÉ A MOINS DE 20%

N.B.: C = CONFORME NC = NON CONFORME

M.E.S.

RESULTATS STATISTIQUES PAR OPTION

OPTIONS	BASE DE L'ANALYSE	NORME (KG/T) MOYENNE	RESULTATS (KG/T) DÉVERSEMENT MOYENNE	DÉVIATION STD	COEFF. VAR.(%)
1. P.G.	3 USINES 1 AN OPÉRATION	11.0	11.34	2.63	23.2
2. P.P.	1 USINE 6 MOIS OPÉRATION	10.0	8.95	3.25	36.3
3. I.G.	3 USINES 6 MOIS OPÉRATION	8.7	10.52	3.01	28.6

OPTIONS TECHNOLOGIQUES (DBO_5 - 1ÈRE ÉTAPE)

PROCEDE AU BISULFITE - (PATE ET PAPIER JOURNAL)

OPTIONS	AVANTAGES	DESAVANTAGES
1. RÉCUPÉRATION DES LESSIVES USÉES DE CUISSON	- RÉCUPÉRATION D'UNE PARTIE DES PRODUITS CHIMIQUES - RÉCUPÉRATION D'ÉNERGIE	- INSTALLATION TRÈS COÛTEUSE - PROCESSUS TRÈS COMPLEXES ET VARIÉS
2. FABRICATION DE SOUS-PRODUITS		- ANALYSE DE MARCHÉS
3. CHANGEMENT DE PROCÉDÉ	- DIMINUTION IMPORTANTE DE PERTES EN MATIÈRES DISSOUTES - ÉCONOMIE DE BOIS (90% VS 50%) - QUALITÉ DU PRODUIT - ÉCONOMIE DE MAIN D'OEUVRE	- TECHNOLOGIE RELATIVEMENT RÉCENTE - COÛTEUX EN ÉNERGIE

DBO_5 (1ère étape)

Analyse du choix des options

Procede	Options	Avant 1980	Apres 1980		
			Complété	En cours	A venir
- Bisulfite et sulfite	1. Récupération lessives	1 (partiel)	1 (partiel)	1 (partiel)	1
	2. Fabrication sous-produits	1	1	1	0
	3. Changement de procede	3	5	3	5
- Kraft	4. Récupération lessives	10	10	0	0

Note: Depuis 1980, 13 procédés au bisulfite sont ou seront abandonnés

Reductions de DBO_5 (1ère étape)

Rejets (T/D)

Avant 1980:	1 400	
Aujourd'hui:	920	
Reduction:	480	(34%)

Couts (MM$)

Depense:	420*	860
En cours:	125*	
A depenser:	315	

* Subventionné a environ 20%

METHODOLOGIE POUR LA DBO_5 (2E ETAPE)

TRAITEMENT BIOLOGIQUE AEROBIQUE

	ETANGS AERES	BOUES ACTIVEES
AVANTAGES	DÉTOXIFIE L'EFFLUENT RÉDUIT LA DBO_5 PEU SENSIBLE AUX VARIATIONS TRÈS PEU DE BOUES CONTRÔLE DES ELÉMENTS NUTRITIFS PEU DE SURVEILLANCE	FAIBLE ESPACE REQUIS RÉDUIT LA DBO_5 RAPIDEMENT
DESAVANTAGES	GRAND ESPACE REQUIS CONSOMME DE L'ÉNERGIE	NE DÉTOXIFIE PAS L'EFFLUENT SURVEILLANCE ÉTROITE SENSIBLE AUX VARIATIONS GRAND VOLUME DE BOUES A DISPOSER

METHODOLOGIE POUR LA DBO_5 (2E ETAPE)

TRAITEMENT BIOLOGIQUE ANAEROBIQUE

	AVANTAGES	DESAVANTAGES
REACTEUR (CELLULE)	FORTES CHARGES: 600 MG/L FORTE RÉDUCTION DE DBO_5 RAPIDE, ÉCONOMIQUE GÉNÈRE UN GAZ COMBUSTIBLE	SÉGRÉGATION DES EFFLUENTS PEU UTILISÉ EN PÂTES ET PAPIERS DOIT ÊTRE SUIVI D'UN AÉROBIE NE DÉTOXIFIE PAS

COÛT ESTIMÉ DES INSTALLATIONS REQUISES (MM$): 620

CRITERES SPECIFIQUES DE CONCEPTION DES ETANGS AERES

TEMPS DE SEJOUR:	5 - 8 JOURS
TAUX DE CHARGE:	0.02 - 0.05 KG DBO_5/M^3- D
PUISSANCE AERATION:	1 - 3 WATTS/M^3
TAUX DE TRANSFERT OXYGENE:	1.5 - 2.0 KGU_2/KWHRE
RAPPORT DBO_5/H.P.:	20 - 40 KG DBO_5/HP
AGENTS NUTRITIFS:	DBO_5/N/P : 100/0.8/0.4
pH :	6.5 - 8.5

POLLUANTS NON CONVENTIONNELS ET TOXIQUES

OBJECTIFS MENVIQ

- REDUIRE LA CONCENTRATION OU L'EFFET D'UN POLLUANT A UN NIVEAU ACCEPTABLE DANS LE MILIEU AQUATIQUE AFIN DE SATISFAIRE LE CRITERE DE QUALITE EN FONCTION DES USAGES
- TRAITER, SUBSTITUER OU ELIMINER A LA SOURCE

CONTRAINTES

- LIMITE DES PROCEDES TECHNOLOGIQUES DE TRAITEMENT
- ASPECT ECONOMIQUE
- FIABILITE DES METHODES ANALYTIQUES
- LIMITE DES CONNAISSANCES

METHODOLOGIE

- TRAITEMENT SECONDAIRE OU L'EQUIVALENT

APPROCHE UTILISEE POUR LA DEFINITION DES OBJECTIFS

I- DETERMINATION DES CONCENTRATIONS LIMITES

- ANALYSE DES COURS D'EAU (NON EXHAUSTIF)
- DETERMINATION DES PROFILS D'ECOULEMENT
- INVENTAIRE DES USAGES
- ECHANTILLONNAGE DES EFFLUENTS
- REVUE DE LA LITTERATURE
- DEFINITION DE CRITERES ET LIMITES
- REVUE DES TECHNIQUES DE DEPOLLUTION

APPROCHE UTILISEE POUR LA DEFINITION DES OBJECTIFS

II- APPLICATION DU CONCEPT DE LA ZONE DE MELANGE

- ZONE AU-DELA DE LAQUELLE LES OBJECTIFS DE CONCENTRATION SONT SATISFAITS
- PETIT OU MOYEN COURS:
  - HYPOTHESE DU MELANGE PARFAIT DES EFFLUENTS SUR LA MOITIE DE LA LARGEUR DE LA RIVIERE SI LE MELANGE SE FAIT A L'INTERIEUR DE 300 METRES
- GRAND COURS D'EAU:
  - CONE DE DIFFUSION DE 300 METRES DE LONGUEUR OU 50 METRES DE LARGEUR

CALCUL DE LA CONCENTRATION TOLERABLE

$$C_E = \frac{(C_R - C_A) \cdot V \cdot L \cdot \tan \theta \cdot H}{Q_E}$$

C_E = CONCENTRATION TOLERABLE A L'EFFLUENT (GRAND COURS D'EAU)

C_R = CONCENTRATION VISEE ET CHRONIQUE (OBJECTIF)

C_A = CONCENTRATION EN AMONT

Q_E = DEBIT DE L'EFFLUENT

θ = ANGLE DU CONE DE DIFFUSION

L = LONGUEUR DU CONE DE DIFFUSION (300 METRES)

V.H = VITESSE ET PROFONDEUR D'EAU BASEES SUR ETIAGE 2 ANS/7 JOURS

VALEURS TYPIQUES DE CONCENTRATIONS

$$C_R = \frac{\text{CONCENTRATION AIGUE}}{45} \text{ (LORSQUE DONNÉES ABSENTES)}$$

CONCENTRATIONS AIGUES		MG/L
ACIDES RESINIQUES		0.2 - 0.7
ACIDES GRAS		12.0
SUBSTANCES PHENOLIQUES, O. M. P-CRESOL:		0.12 - 3.5 - 0.4
DERIVES CHLORES	- ACIDES RESINIQUES:	0.6
	- ACIDES GRAS:	2.5
	- SUBS. PHENOL:	0.3 - 0.7
BPC, AROCLOR (LIMITE DE DETECTION):		0.01 µG/L
HCB (LIMITE DE DETECTION):		0.001 µG/L
REACTIFS:	- CL_2	2 µG/L
	- XYLENE:	10

CHARGE ACCEPTABLE

MEILLEURE TECHNOLOGIE PRATIQUE (BPT)

OBSERVATIONS GENERALES SUR LES TRAITEMENTS CONJOINTS

RESERVES SUR:

- PROLIFERATION DES BACTERIES PATHOGENES
- EFFICACITE GLOBALE DU SYSTEME
  - TEMPERATURE D'OPERATION
  - DILUTION
  - DEGRADATION BIOLOGIQUE PREFERENTIELLE
  - INHIBITION BACTERIENNE
- PRODUCTION DE COMPOSES CHIMIQUES INDESIRABLES DU A LA PRESENCE RESIDUELLE DE REACTIFS CHIMIQUES
  - AU TRAITEMENT
  - DANS L'EMISSAIRE
- CONCEPTION DES SYSTEMES
  - PRETRAITEMENT
  - MODULATION
  - FLUCTUATION DE CHARGES
  - FLUCTUATION DE DEBITS

COMPARAISON POLLUTION URBAINE ET PATES ET PAPIERS

- GLOBAL

PARAMETRES	URBAIN	PATES ET PAPIERS
DEBIT (M^3/D)	2.75×10^6	2.5×10^6
DBO_5 (T/D)	380	1 400

- SITUATION TYPIQUE (USINE DANS UNE VILLE MOYENNE)

	VILLE	USINE	RAPPORT
DEBIT (M^3/D)	3 000	60 000	20
DBO_5 (T/D)	0.5	25	50

RECENT DEVELOPMENTS IN ELECTROCHEMICAL TREATMENT OF WASTEWATERS

M.I. Ismail[1], Y.M.Bakir[2] and M.M.Bakir[2]

1. University of Waterloo, Waterloo, Ontario, Canada
2. Alexandria, Egypt

ABSTRACT

The up-to-date state of the art of electrochemical treatment of wastewaters is summarized. Major techniques can be broadly classified as static, flow through porous electrochemical reactors of various types. The selection of the technique is dependent on the type of the pollutant and the type of the wastewater to be treated.

The electrochemical techniques are widely used for treatment of wastewaters from the electroplating, electroforming, and other similar industries. The removal of heavy metals, in some cases, offsets the cost of the electrochemical reactor used for wastewater treatment. Several types of the commercially available electrochemical reactors are discussed.

Introduction

This report reviews the state-of-the art for wastewater treatment using electrochemical processes. The reader is assumed to be familiar with the basic principles in electrochemistry, wastewater terminology and water pollution aspects of the various processes.

Monitoring Considerations:

Monitoring of wastewater treatment processes requires considerable attention by management. The process chemistry for water and wastewater treatment has been recently reviewed[1-3]. The simple electrochemical reaction that occurs when zinc and copper are connected together in a galvanic cell type is shown in Fig. 1.

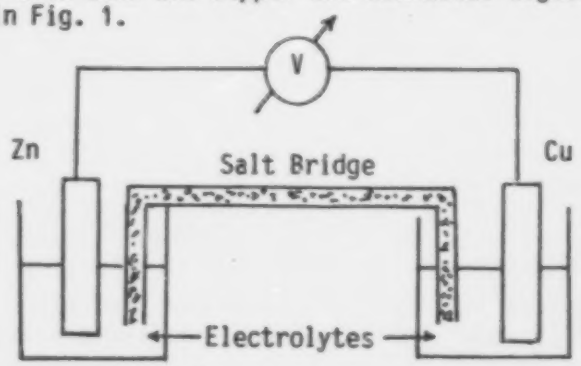


FIG. 1 GALVANIC CELL Cu AND Zn

The reaction represented as

$$Zn_{(s)} + Cu^{2-} : SO_4^{2-} \longrightarrow Cu_{(s)} + Zn^{2+} : SO_4^{2-}{}_{(aq.)} \quad \ldots\ldots(1)$$

The zinc looses 2 electrons and is oxidized to Zn^{2+}, while Cu^{2+} gains two electrons and is reduced to elemental Cu. This is a typical galvanic cell which needs no external power. If the two electrodes (e.g., Zn and Cu) are the same, an external power source is needed to affect the electrochemical reaction (oxidation of one and reduction of the other electrode). The correlation between the electrochemical half cell (half reaction) potential and cell potential is governed by Nernst equation

$$E_{cell} = E^{\circ}_{cell} - (RT/nF) \ln Q$$

where n = number of moles of electrons transfered as indicated in the balanced chemical equation,

F = Farady Constant = 23061 Cal/mole-volt
E_{cell} = Potential of a Cell (volts)
E°_{cell} = Potential of a Cell at Standard Conditions
Q = Quotient

$$= \frac{\text{Oxidized form}}{\text{Reduced form}} = \frac{[Zn^{2+}]}{[Cu^{2+}]}$$

The relationship between the standard cell potential and the free energy change is given by the equation

$$\Delta G^{\circ}_{reaction} = -nF E^{\circ}_{cell} \quad \ldots\ldots\ldots\ldots\ldots\ldots(2)$$

If E°_{cell} is negative, $\Delta G^{\circ}_{reaction}$ will be positive and the reaction will not be thermodynamically feasible.

Identification of Contaminants

More than 100 specific organic compounds have been identified in certain wastewaters [4] using GC/MS for the analysis for limits higher than 0.1 µg/l. Over 2000 papers are available in the literature during the last decade related to wastewaters applied to main 6 industries.

1. Pulp and Paper
2. Metal Finishing
3. Heavy Chemicals (inorganics)
4. Petrochemicals
5. Metallurgical Industries (iron and steel)
6. Food products and Agriculture Industries
7. Groundwater

Various techniques are utilized for wastewater treatment. The electrochemical techniques are of interest for the electrical products industries and metal finishing for both ferrous and non-ferrous types. Some industries are more suited to the electrochemical techniques than others. For example, the photo processing industry uses electrolytic recovery for Ag and for other precious

metals. The control and measurements of organic pollutants in wastewater are dependent on the lower hazard limit of such specific compounds.

The U.S. Environmental Protection Agency (EPA) identified 129 priority pollutants with goal of establishing industrial effluent limitations and guidlines [5]. There is no problem for metal detection at the level of 25 ppb but the organic contaminants represent the main problem due in part to thermal decomposition to other compounds, oxidation or poor chromatographic properties.

Continuous monitoring instrumentation (on-line) allows greater flexibility for treatment and control. High pressure liquid chromatography (HPLC) is an example of instrumentation for determination of organics. The various monitoring systems may use the variation of the physical/chemical characteristics of the water as a result of the presence of contaminants. The dielectric constant, (D.E.) or permittivity of water is 80 and is one of the highest known showing little variation up to frequencies of about 80 MHz [6]. The dielectric value is reduced by organic contaminants [7]. The conductivity of pure water is 4.2×10^{-6} mho/m but this is increased by soluble salts. The modern analytical techniques for water quality measurements are available for organic and inorganic water pollutants [8]. Pulse polarography is now well understood as a technique and could be applied to water quality problems. Stripping voltametry and ion selective electrodes are convenient for routine analytical determination of many ions in water solutions. Several sensors are now available as shown in Table 1.

TABLE 1. SELECTED SENSORS FOR WASTEWATER TREATMENT MONITORING

Type	Range	Selected Manufacturers
Chloride	10 - 1000 ppm	Great Lakes
pH/Electrode	2 - 12 pH/units	Orion
Sodium Electrode	10 -1000 ppm	Beckman
Hardness/Electrodes	1 - 1000 mg/l	Orion
Total Residual Chlorine/Electrode	0 - 1000 ppm	Orion
Conductivity/Impedence Bridge	0 - 2000 μmho/cm	Beckman
Dissolved Oxygen/Electrode	0 - 20 mg/l	Delta Scienfific

Water Desalination By Electrodialysis

Desalination of water is done commercially by two main processes, electrodialysis and reverse osmosis. The principle of electrodialysis is that the wastewater solution (rich in metal ions) is subjected to an electric field (D.C.) between two electrodes with a continuous potential difference applied between them. Cations will be attracted to the cathode and the anions attracted to anode. If several membrances are placed between the two electrodes and these membranes are permeable to certain ions (i.e., negative membranes are permeable to cations and positive membranes are permeable to anions). Ions are restricted in migration as shown in Fig. 2. One of the largest electrodialysis units is used in Benghazi (Libya) (1969). It produces 800 m^3/h of water of 500 mg/l total dissolved solids from water containing 2000 mg/l. This unit incorporates 8 units of 100 m^3/h operating in parallel. Other countries have similar units. The cost of produced

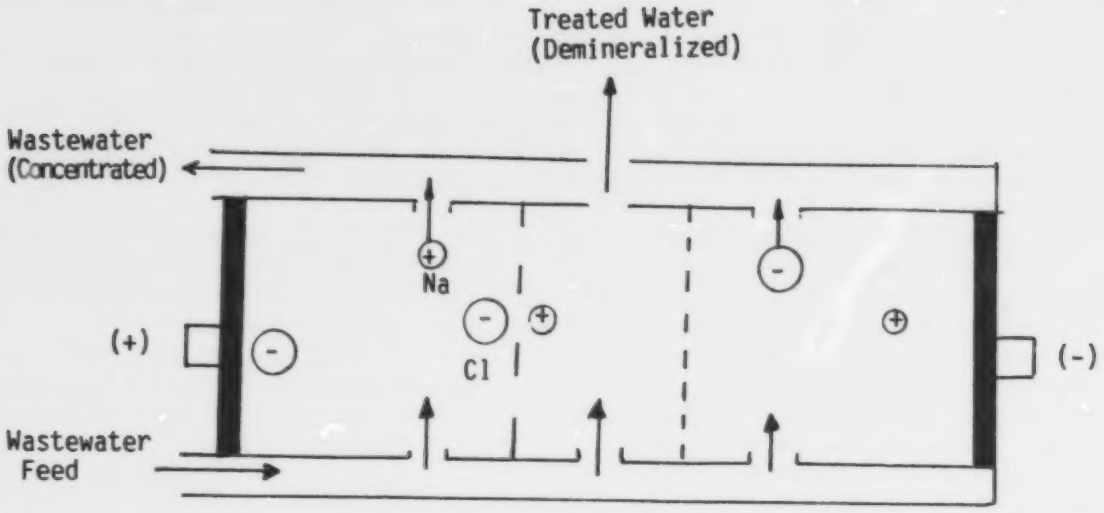


FIG. 2 BASIC SEPARATION PRINCIPLES : ELECTRODIALYSIS

desalined water increases with the increase in salt content of the feed water. Water of 3 g/l can easily be treated by this method to produce potable water. The energy consumption sould be in the order of 40 KWh/m^3 produced.

With this process the water is divided into 2 fractions ; small concentrated saline solution and a partially demineralized water, (Fig. 2). This process is suitable only for slightly brackish water. The capacity of membranes falls as the salinity of water increase. It is economic with waters of salinity≤3 g/l. The amount of electricity increases as the salinity of water decreases (resistivity of water increases). This method is of little use with water of salt less than 0.5 g/l. The electrodialysis process is appropriate for daily flows up to several thousands of cubic meters. The desalination process is usually carried out in several continuous stages in series as shown in Fig. 3.

Advantages and limitations of electrically driven membrane processes electrodialysis and transport depletion processes for wastewater (Desalination of brackish and sea water) have been reviewed [9-12]. Most dissolved solids in brackish waters are inorganic in nature. The various types of membrane processes utilized in wastewater treatment are shown in Table 2 and Fig. 4. The chosen system must show variations in conductivity, potential (e.g., Potentionetry, Amperometry, Coulometry, Chronopotentiometry, or Voltametry [13].

several reviews on porous electrode systems for water treatment are available for such new reactors [14-16]. Fluidized bed electrolytic cells could be used for the removal of toxic heavy metal ions from solutions [17]. The removal of heavy metal ions from dilute solutions in electroplating-type effluent may result in considerable savings [18]. Industrial reactors for environmental polution control are commercially available [14-16, 19]. Several metallic ions such as cadimum, lead, chromium can be removed as shown in Table 3.

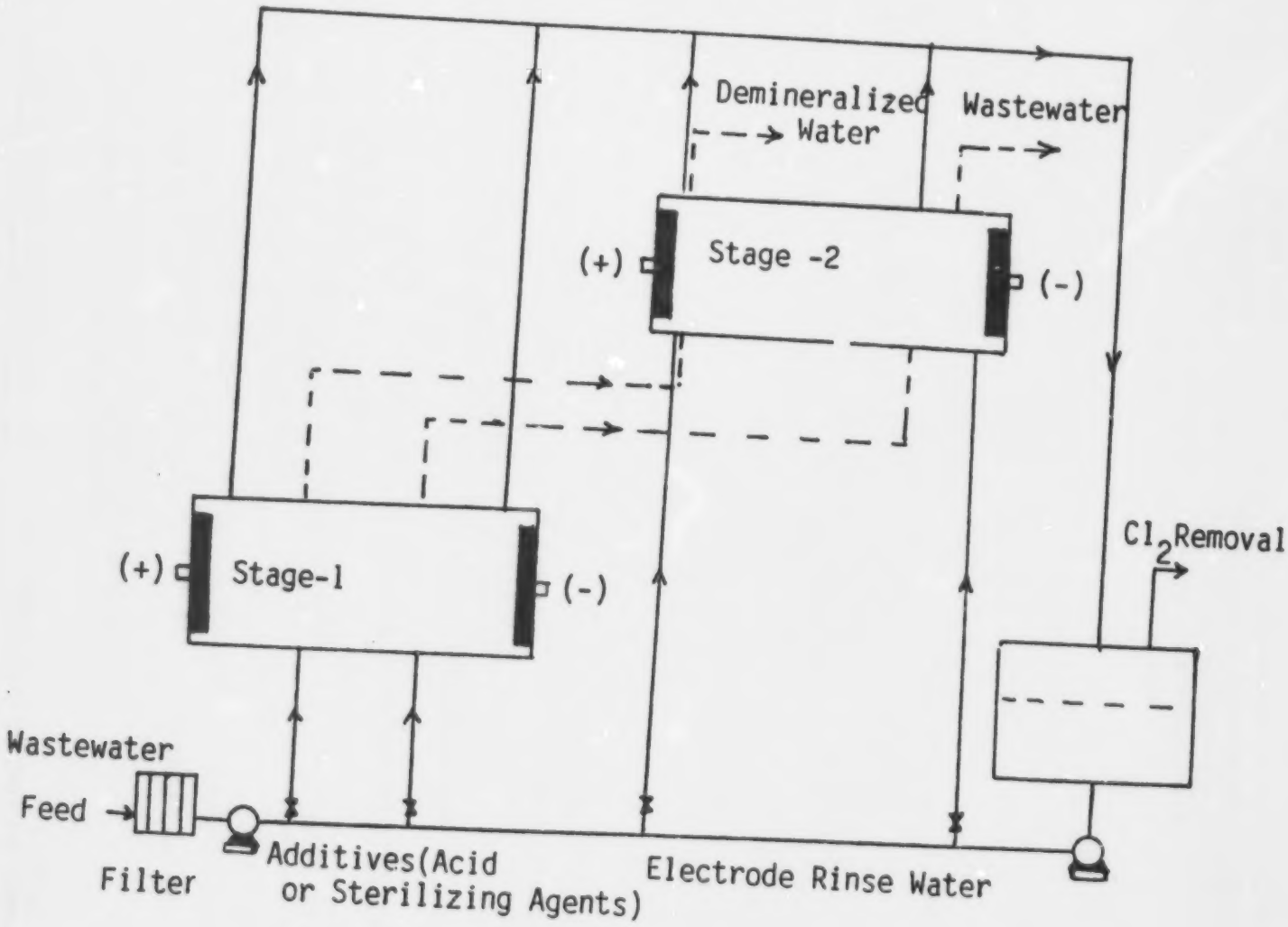


FIG. 3 CONTINUOUS SYSTEM : MULTI-STAGE ELECTRODIALYSIS

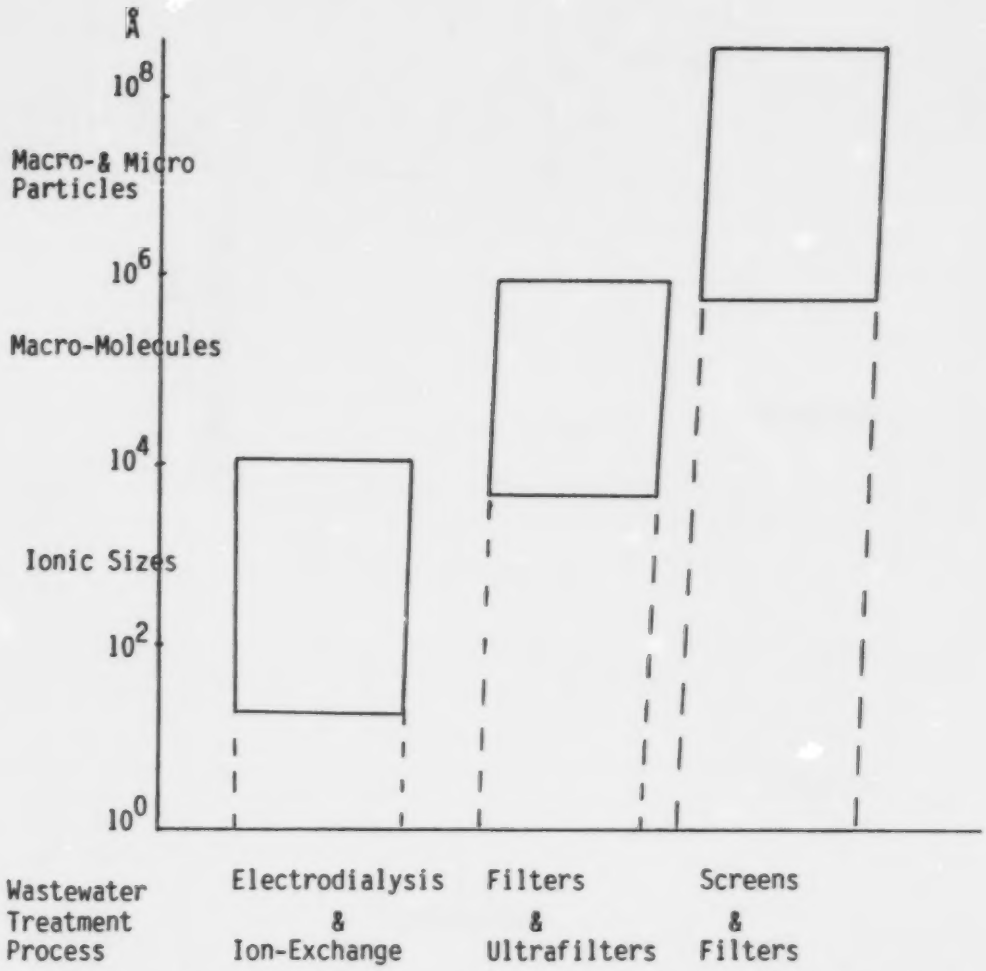


FIG. 4 VARIOUS MEMBRANE PROCESSES UTILIZED IN WASTEWATER TREATMENT

TABLE 2. ELECTRICALLY DRIVEN MEMBRANE PROCESSES FOR WASTEWATER TREATMENT

Process Property	Electrodialysis	Electro Depletion
Driving Force	Electrical Potential	Electrical Potential Difference
Pollutants Removed	Soluble Inorganics	Soluble Inorganics
Remaining Constituents	Little Salt	Salt (more than in electrodialysis)
	All Organics including (Viruses, Bacteria)	All Organics Soluble or Non Soluble Including Bacteria
Size Range of Permeable Species, A°	3 - 300	10 - 1000

Electrodialysis and Packed Beds

As the concentration of ions drops, the resistance of the solution increases. The presence of the resin bed maintains the conductivity even at low ion concentrations(on the levels of a few ppm). Fig.5 shows the basic principle of electrodialysis for production of acid/alkali from salt solution using a resin bed as an example of the system.

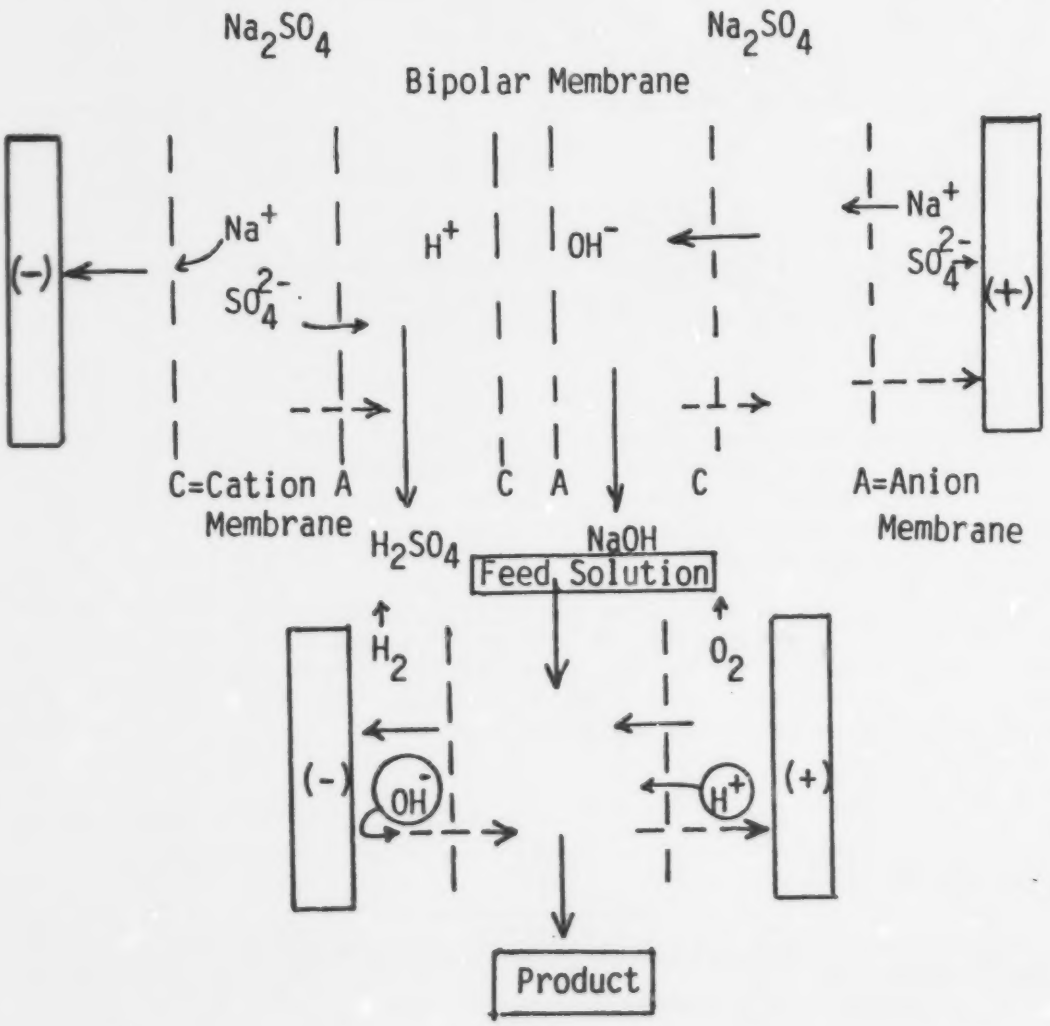


FIG. 5 PRODUCTION OF ACID/ALKALI BY ELECTRODIALYSIS OF SALT SOLUTION

TABLE 3. TYPICAL CHANGE IN WASTEWATER CHARACTERISTICS BEFORE AND AFTER ELECTROCHEMICAL TREATMENT

Contaminant	Before mg/l	After mg/l
Copper	0.1	0.1
Nickel	2.3	0.1
Chrome	2.1	0.1
Cadmium	0.3	0.1
Zinc	2.7	0.3
Lead	0.1	0.1
Cyanide	7.5	0.1

Electroosmosis is used for dewatering of sludges. An electric field applied between two electrodes causing preferential movement of water towards the cathode and thus water drainage from sludge occurs (90 % H_2O) at low pressure [6].

Future Trends

Metal removal/recovery from dilute solutions should meet environmental regulations and process economics should be compatible with other non-electrochemical methods for commercialization. Low cost electricity is an advantage for electrochemical methods particularyly during the off peak times. Also, the electrochemical processes are faster than most of other waste treatment systems.

The wastewater solutions are fed to the reactors either under the effect of gravity or utilizing suitable pumping system. The flow rates in laboratory experiments range from 10 - 10^3 cm^3/min depending on the type and cross sectional area of the reactor. Fig. 6 shows schematic diagrams of some reactors of potential interest for wastewater treatment.

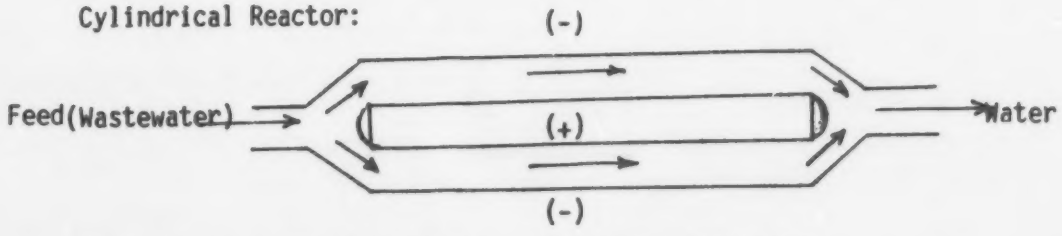
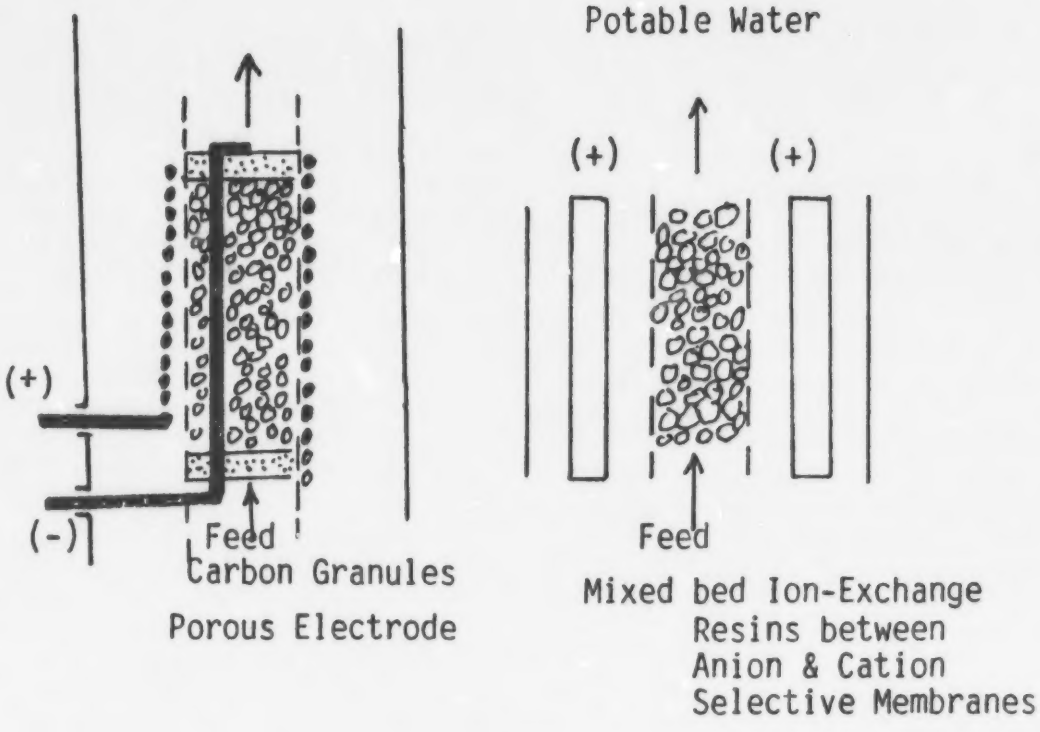
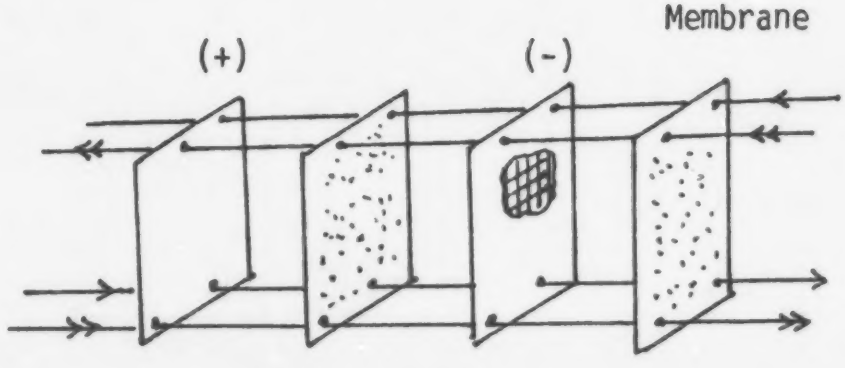


FIG. 6a SELECTED REACTORS FOR WASTEWATER TREATMENTS





Filterpress-type Electrochemical Reactor
for Wastewater treatment

FIG. 6b SELECTED REACTORS FOR WASTEWATER TREATMENTS

Lower Cost Electrode Systems

Granular carbon electrodes are low cost and convenient with respect to other conducting particles of Ni, Cu, Ah, Pb, etc. An average diameter of 1 mm which can be consolidated into compact highly porous electrodes for medium conductance solutions and current density. The potential set up for electrolysis will be constant. The potential of the porous electrode is always cathodic so that heavy metal ions such as Cd, Hg, Cu, and Pb deposit and electroplate on it. In a continuous flowing system the current achieved depends on the flow rate and on the applied potential. For a single pass of solution through the packed bed electrode at steady state, cathodic current is given by the equation

$$I = ZFJC\,(1 - \exp - h/\lambda) \qquad \ldots\ldots\ldots\ldots\ldots \quad (3)$$

where I = Cathodic Current
ZF = the Molar Ionic Charge
J = Flow Rate
C = Inlet Concentration of Electrodepositing Ions
h = Height or Length of the Porous Bed

$$\lambda = J/ADa\epsilon \simeq 1 - 10 \text{ cm} \qquad \ldots\ldots\ldots\ldots\ldots \quad (4)$$

λ = Characteristic Length Defined by Equation (4)

where A = Cross Section Area of the Bed
a = the Specific Area of the Cargon (cm^2/cm^3)
ϵ = Voidage or Gross Porosity of the Bed

The current increases non-linearly with J and with h. The outlet or effluent concentration, C_h can be calculated from the equation :

$$C_h = C \exp -h/\lambda \qquad \ldots\ldots\ldots\ldots\ldots\ldots \quad (5)$$

and the efficiency of the bed in removing ions by electroplating

$$= \frac{C - C_l}{C} = 1 - \exp -h/\lambda \qquad \ldots\ldots\ldots \quad (6)$$

The heavy metal removal efficiency increases with h to a maximum value of one and decreases with increasing J to a minimum value of Zero as shown in Fig.6.

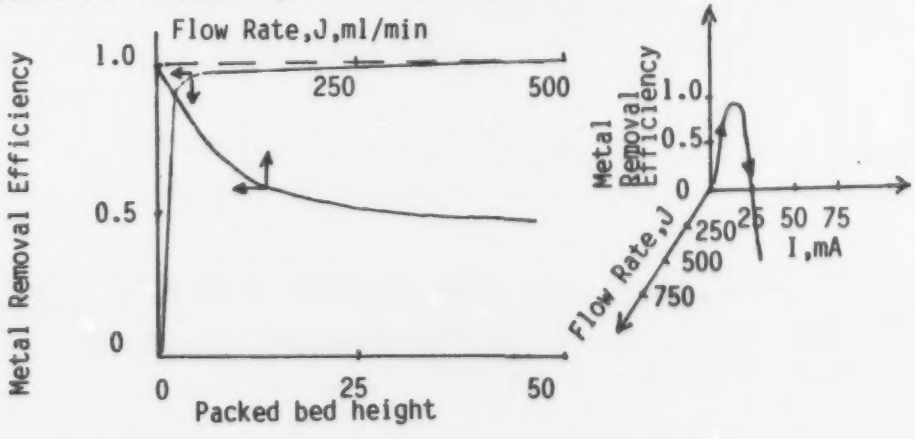


FIG. 7 METAL REMOVAL EFFICIENCY RELATED HIGHT OR LENGTH OF POROUS BED

For a bed of 25 cm, and a bed diameter of 5 cm and flow rates of less than 250 cm^3 /min the efficiency of removal will be less than 95 % [16] and in practice it is even more than 99 %. The metal removal efficiency is independent of the initial concentration of ions but on the height of the packed bed or on recirculation using the same height of bed. At concentrations 10^{-3} to 10^{-4} M, the current efficiency is almost 100 %. The progress of metal deposition on porous bed electrode will result in welding and sealing the pores which reduces bed activity unless the current (polarity) is reversed to remove the precipitated metals (the process known as anodic stripping). This process is not mass transfer controlled and is therefore independent of the flow rate. The recovery of the metal is dependent on the stripping time and is almost 100 % if sufficient time is used a few minutes in commercial reactors). In the stripping process the volume is limited so the concentration of metal is very high (10^3 M) which could be useful for some cathodic processes such as electroplating and surface finish processes which in some cases this may pay the cost of the wastewater treatment process for Ag, Au and other expensive metals.

For the removal of dissolved anions which can be anodically trapped as insoluble film on reactive metal (e.g. Cl on Ag or photosphate on Zn). The regeneration could take place cathodically and any leakage of the metal out of its anodic reactive state can be trapped by using a second following bed as the cathode as shown in Fig. 8.

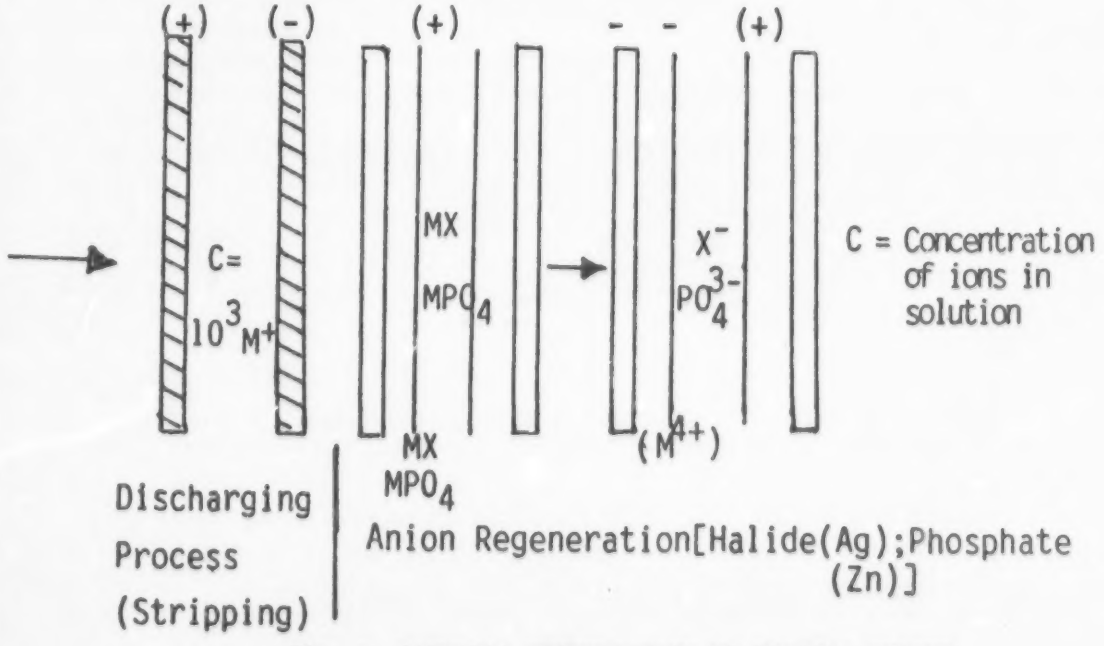


FIG. 8 CATHODIC REGENERATION OF TRAPPED ANIONS

Commercial power supplies with adjustable current reversal ratios are of potential interest for wastewater electrochemical reactors [20].

Carbon Fibers and Graphite Electrodes

Graphite bed always recovers its activity when it is left for a while without flowing of the wastewater solution. However, the used bed will not regain its initial efficiency for metal removal or ion desorption processes. Carbon fiber electrodes are used commercially for wastewater treatment.

The relative merits of electrochemical wastewater treatment processes are shown in Table 4.

TABLE 4. RELATIVE MERITS OF ELECTROCHEMICAL WASTEWATER PROCESSES

The process	Advantages	Disadvantages
-Electrochemical Reactors	-For organic oxidation	-not in practice as final step
-Electrochemical Reactors	-Highly active electrode	
-High Surface Area Reactors	-highly active electrode per unit volume	-Premature fouling and short circuiting
-Rotary cylinder electrode	-Compact	-Maintenance cost
-Filter-press Reactor	-Compact	-Relatively high pressure drops
-Rotating wiper-blade electrode Reactor	-Compact, good separation	-High construction, maintenance operating costs

The relative cost of various wastewater treatment processes are shown in Fig.9[21].

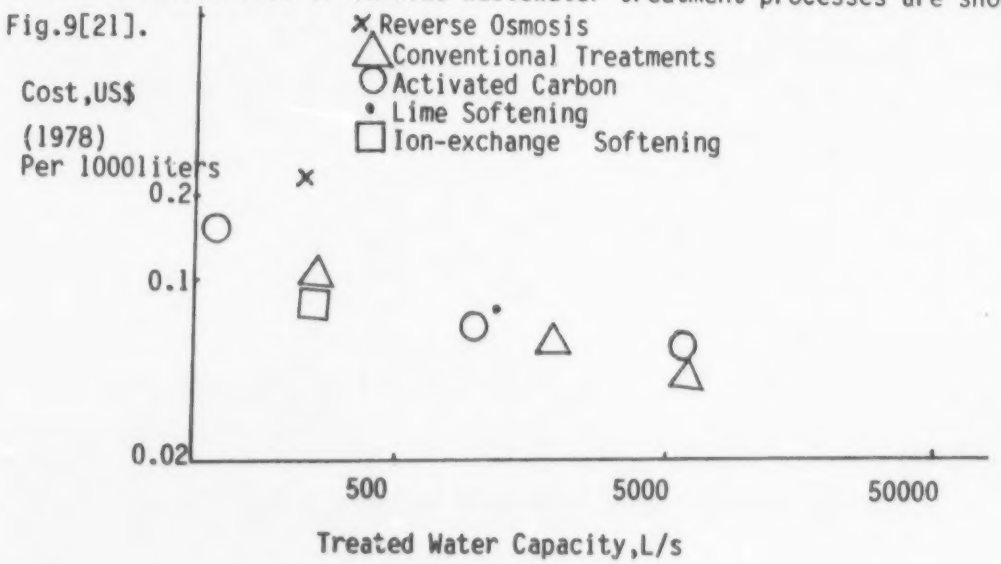


FIG. 9 COST OF WATER TREATMENT (log-log Scale)

Recent trends in electrochemical wastewater treatment are summarized in Table 5. Periodic current reversal and pulsed processes are gaining more use.

TABLE 5. RECENT TRENDS IN ELECTROCHEMICAL WASTEWATER

Process Trends	Remarks	Reference
-Reversed current	-Pulses causes electrode activation and descaling	22
-Ultrasonic	-Descale of membrane	
-Silicon carbide coating	-Protection of graphite anode	
-Electro coagulation	-Al anode	
-H_2S, gas removal	-Reversed current	
-Dissolved pollutants	-Active filters, porous electrodes	19
-Electroplating Effluents	-Electrodeposition	23
-Seawater	-Production of Salt	24
-Sewage water	-Metal removal	25

Conclusions

Electrochemical techniques are promissing for various wastewater treatment processes. Their cost is comparable to other processes. Periodic current reversal and pulsed processes are gaining more potential in electrochemical treatments of wastewater.

Acknowledgement

Valuable discussions with Professor T.Prasad and Professor G.J.Farquhar are greatly appreciated.

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DESIGN AND OPERATION OF AN ACTIVATED BIOFILTER (ABF) PLANT AT SAINT JOHN, NEW BRUNSWICK

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ABSTRACT

The activated biofilter (ABF) process is a hybrid of attached (trickling filter) and suspended (activated sludge) growth systems. An activated biofilter plant of 6820 m^3/d capacity was recently built in Millidgeville, Saint John, New Brunswick and became operational by the end of 1983. It is one of the few plants built in Canada employing the ABF process.

This paper presents the basic process concepts and the process design aspects of ABF systems, along with a review of U.S. experience of these systems, especially the performance of the ABF plant in the City of Helena, Montana. This paper discusses the process design and operational features of the ABF plant recently built in Millidgeville, Saint John and provides preliminary results of its performance.

INTRODUCTION - BASIC PROCESS CONCEPTS

The activated biofilter (ABF) process is a hybrid of attached (trickling filter) and suspended (activated sludge) growth systems. Neptune Microfloc markets the ABF process and the horizontal redwood biomedia used for the biological treatment of wastewater. The four major components of the ABF process are a redwood media tower filter called a biocell, a biocell recirculation system, a short-term activated sludge aeration tank and a conventional secondary clarifier. Figure 1 shows the process flow schematic of an ABF system. The flow to the biocell consists of a combination of raw wastewater, underflow from the biocell and recycled sludge from the secondary clarifier. This stream is a mixed liquor with a relatively high suspended solids concentration. A fixed-film growth occurs on the media in the biocell. The mixed liquor that circulates over the media contains biologically active organisms very similar to those found in activated sludge systems. The ABF system is thus a combination of the activated sludge and the trickling filter systems.

Recirculation of activated sludge to the biocell is found to be important since it serves to: (1) substantially increase the effective BOD removal in the biocell, by adding to the quantity of organisms in contact with the wastewater as it passes through; and (2) contribute to the health and viability of the mixed liquor, and therefore, to overall process stability. Pilot studies have demonstrated the superior stability of the ABF process over the similar roughing filter/activated sludge process.

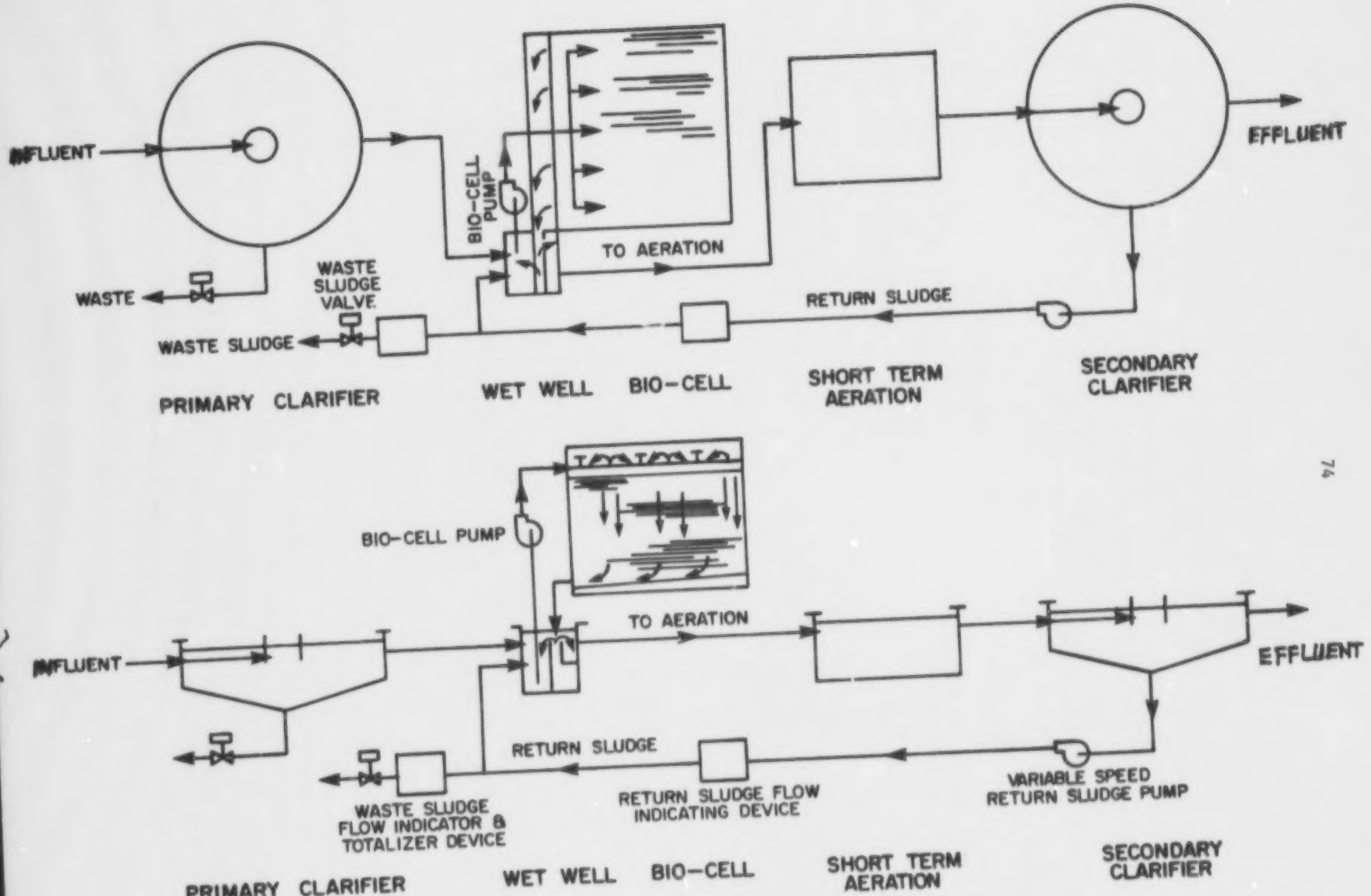

FIGURE I. ACTIVATED BIOFILTER PROCESS FLOW SCHEMATIC

The biomedia used in the ABF system consists of horizontal redwood slats with the high specific surface (460 m^2/m^3). The rough sawn texture of the redwood facilitates retention of biological film. Oxygen transfer is achieved by the dual action of the mixed liquor continually moving in a film across the biota and splashing between the wood slats. The serpentine pattern of flow through the media in a horizontally-packed filter significantly increases the detention time of the liquid as it passes through the system. The depth of media is typically 4.27 m.

PROCESS DESIGN ASPECTS

The ABF process design is based on the results of various pilot and full scale plant evaluations. Slechta and Mattli[2] found that the ABF process could provide a high quality effluent and a settleable sludge at very high process loadings. Their analysis led them to define the organic loading to the process in terms of a system F/M expressed as follows:

$$\text{System F/M} = \frac{\text{Primary effluent kg BOD/day to biocell}}{\text{kg MLVSS in aeration basin only}}$$

This expression does not account for the mass of fixed growth microorganisms in the biocell which cannot be quantified. The System F/M is therefore several times greater than that used in activated sludge design. Slechta and Mattli determined that the limiting design constraint for the process was the aeration basin oxygen demand and not System organic loading in which:

$$\text{Aeration Basin Oxygen Demand} = \frac{\text{kg } O_2 \text{ to aeration basin}}{\text{kg BODR in System}}$$

For most domestic wastes, the biocell is typically designed at a BOD loading of 3.2 $kg/m^3.d$ which falls well within demonstrated operating ranges and allows for widely varying load fluctuations. Research by Owen and Slechta[3] and design models for fixed film reactors predict an approximate BOD removal of 65% at this loading. The biocell is therefore designed as an efficient roughing unit which reduces the loading on the downstream aeration basin. Its effect on the overall process is to reduce the aeration basin oxygen demand to about 0.37 kg/O_2 per kg BOD removed in the process. This demand is approximately 30 to 40% of that which would be required by an activated sludge plant.

Slechta and Mattli[2] reported that there is no correlation between ABF process performance and biocell height over a range of 2.1 to 6.7 m. CH2M-Hill[4] found that the performance of a fixed film reactor is related to organic volumetric loading and independent of biocell height above 2.4 m. Cook[5] concluded that at a given organic volumetric loading, there is very little increased COD removal above 1.2 m height. These studies indicated that the height of the biocell can vary substantially without affecting ABF process performance.

Typical ABF design parameters for carbonaceous and nitrification systems generally recommended by Neptune Microfloc[6,7] are presented in Table 1. Figure 2 shows the variation in aeration basin oxygen demand in relation to

Table 1.

TYPICAL ABF DESIGN PARAMETERS[6,7]

Criteria and Parameters	Units	Value	Typical Range
Effluent Criteria			
BOD_5	mg/L	20	10-30
Suspended Solids	mg/L	20	10-30
NH_3-N (Nitrification Design Only)	mg/L	1.0	0.5-2.5
Biocell Parameters			
Organic Load	$kg/m^3.d$	3.2	1.6-5.6
Media Depth	m	3.27	2.14-6.71
Hydraulic Parameters			
Sludge Recycle	fraction of raw flow rate	0.50	0.3-1.00
Biocell Hydraulic Load (minimum)	$L/M^2.S$	0.82/1.23[1]	0.82-4.48
Biocell Recycle-as required to maintain media wetting rates or as dictated by diurnal flows			
Aeration Parameters			
System F/M^2			
Carbonaceous Design	kg BOD/day/kg VSS	1.4	0.8-3.5
Nitrification	kg BOD/day/kg VSS	0.3	0.2-0.6
Oxygen Utilization			
Carbonaceous Design	kg O_2/lb BOD rem.	0.37	0.2-0.8
Nitrification[3]	kg O_2/lb BOD rem.	0.65	—
MLVSS Concentration	mg/L	3,000	1,500-4,000
MLSS Concentration	mg/L	4,000	2,000-5,000
Clarifier Parameters			
Overflow Rate	$m^3/m^2.d$	29	15-59
Solids Loading	$kg/m^2.d$	147	59-234

[1] 0.82 $L/m^2.S$ for rotary distributors.
1.23 $L/m^2.S$ for fixed nozzle distributors.

[2] Based on primary effluent BOD_5 (process influent) loading to MLVSS in aeration basin.

[3] Total O_2 required = Carbonaceous + 4.6 kg O_2/kg NH_3-N removed.

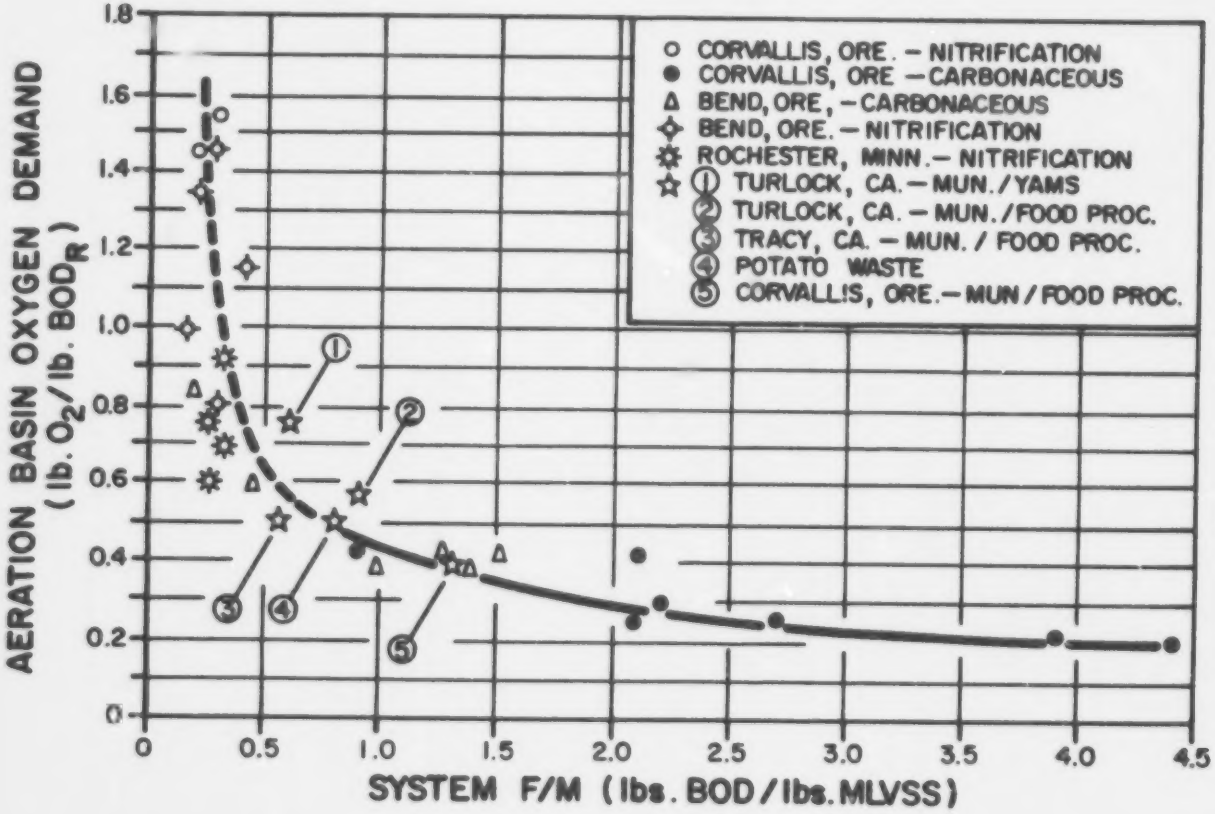

FIG. 2. AERATION BASIN OXYGEN DEMAND VS. SYSTEM F/M

system F/M. It may be observed that the oxygen demand decreases as the system F/M increases.

In the ABF process, the settleability of the mixed liquor obtained from the biocell tower is quite good even under severe shock loading. The settling characteristics of the mixed liquor are very similar to those for trickling filter sludge. Apparently, the slough of biological mass from the filter media combines with the activated sludge suspended in the mixed liquor in such a way as to produce a significantly better settling mixed liquor than for a comparable air-activated sludge system. The ABF system consumes less power than a comparable complete-mix activated sludge system with 5 hours aeration time. The ABF system would consume 60 - 70% of the power that a comparable extended aeration system would consume. The ABF system combines the operational stability of the trickling filter system with the relative high BOD removal efficiency of the activated sludge system. The ABF system also provides reserve capacity which is not available in an extended aeration system or in a conventional activated sludge system.

The heat loss through an ABF system utilizing short-term aeration would be less than the heat loss for an extended aeration system. Heat loss through the filter is low and since aeration time is kept short for aeration, heat loss through the system is low. If proper wetting rates are maintained on the biomedia surface by recirculation of underflow, cold spots and freezing of the tower or media would not occur even at low temperatures. Temperature loss across the biofilter systems is found to be very low, of the order of one degree centigrade[8]. This stable heat loss characteristic of the system is advantageous in cold climate situations. Performance data from the Idaho Falls, Idaho plant indicate that the ABF process is capable of achieving the desired efficiency under cold climate conditions.

EXPERIENCE IN U.S.A. AND CANADA

As per the ABF product line installation list issued by Neptune Microfloc in September 1980, there were 53 ABF installations in North America; of these, 35 are municipal plants, 14 are industrial plants and 4 are package plants. In addition, there were 23 ABF plants without aeration operating in North America in September 1980; of these are 19 municipal plants and the rest are industrial plants. An ABF system of 5700 m^3/d capacity, without aeration, has been operating in Alberta treating oily wastewater for Shell Oil since 1975. A municipal wastewater treatment plant of 9080 m^3/d capacity employing ABF process is operating in Bathurst, New Brunswick since 1981. A wastewater treatment of 6820 m^3/d capacity using ABF process has recently been constructed in Millidgeville, Saint John, New Brunswick.

The ABF system marketed by Neptune Microfloc was first brought into operation in early 1973. Operational experience has been gained over a period of ten years.

Pilot Plant Experience

Numerous pilot plant studies have been conducted especially by Neptune Microfloc for municipalities and industry[1,9] to demonstrate the effectiveness of the ABF process and determine system design parameters. Research and development studies have been conducted by the company to ascertain the significance of various parameters involved in the process. Studies included process evaluation for a wide spectrum of wastes ranging from usual municipal

wastewater to high strength industrial wastewater. Operation conditions and results of pilot plant studies are summarized by Slechta and Mattli[2]. The pilot studies showed that the system is capable of stable performance under increased process loading rates compared to a high rate trickling filter and it produced an effluent of better quality. The ABF process, in fact, has not been found to operate in accordance with classical formulas developed for fixed growth systems[2].

Operating plant experience

Plant operational data obtained from four ABF installations were summarized by Dunnahoe and Hemphill[10]. The above data showed that even though influent BOD and SS were varying widely, the effluent BOD and SS in all the four plants were below 30 mg/L and 40 mg/L.

A survey of 19 operating ABF plants was carried out by Harrison[11]. Based on 19 plant surveys, the most outstanding feature of this process was found to be the stabilizing effect the biofilters gives to the more sensitive activated sludge system. The process has been generally well accepted by the plant operators. Its ease of operation and added process stability are expected to make the ABF process a leading contender, especially where shock loads or industrial waste are involved.

Performance of ABF System at Helena, Montana[12,13]

A field evaluation study was initiated by U.S. EPA at Helena with the primary objective of developing year-round, full-scale operating and performance data on the ABF process. The experimental programme was conducted in such a way that organic and hydraulic loadings imposed during the winter months approached the manufacturer's recommended design criteria.

Three different process loading regimens were investigated over a 17-month period. These regimens corresponded roughly to 1) one half of the manufacturer's recommended loadings on both the tower and the aeration tank, 2) one half of the recommended loading on the aeration tank and the full recommended loading on the tower and 3) the full recommended loadings on both the tower and aeration basin (winter months). Because of available wastewater flows and plant facility operating constraints, the desired 50 and 100 percent loadings in reality averaged closer to 40 and 80 percent of the manufacturer's recommended design criteria.

Excellent overall treatment performance was observed throughout the study. The average final effluent BOD and SS concentrations ranged from 14 to 24 mg/L and from 10 to 27 mg/L respectively. EPA'S monthly-average, 30 mg/L secondary treatment requirement was exceeded during one 5-week stretch for SS only because of operational procedure. Potential savings were indicated in the energy requirements of the ABF process as compared with those of the conventional activated sludge process.

The study concluded that the ABF process is an attractive, competitive secondary treatment alternative because of its operational stability, performance reliability and energy savings potential. The study recommended that consideration should be given to increasing the detention time of the aeration basin beyond that recommended by Neptune Microfloc, especially if the biocell loadings recommended by Neptune Microfloc are used in the design.

PROCESS DESIGN FEATURES OF MILLIDGEVILLE ABF PLANT

Basic Design Data

Design population	-	12000
Average flow	-	6820 m^3/d
Peak flow	-	17270 m^3/d
BOD	-	1090 kg/d
SS	-	1250 kg/d
Minimum required treatment (BOD and SS removal) %	-	85

Treatment Scheme

The treatment system would consist of preliminary treatment - mechanically cleaned bar screen and aerated grit chamber, primary treatment - clarifier, and secondary treatment (ABF process) with chlorination of the effluent, before discharge to the St. John River. Primary sludge and secondary sludge from the ABF process will be mixed, gravity thickened and dewatered using a belt filter press. The dewatered sludge will be hauled away for disposal. Process design features of the primary and secondary treatment units are given below.

Primary Clarifier

A rectangular tank of 6.10 m wide, 34.75 m long and 2.75 m SWD is proposed to be used.

Overflow rate (average flow)	-	32.2 $m^3/m^2.d.$
Overflow rate (peak flow)	-	81.5 $m^3/m^2.d.$
Detention time	-	2.1 h

Biocell

Assuming 35% BOD removal in primary clarifier, biocell influent BOD	=	709 kg/d
Biomedia volume	=	228.4 m^3
Biocell area	=	53.6 m^2
Depth of media	=	4.27 m
Biocell BOD loading	=	3.1 $kg/m^3.d.$

Biocell Pumping Requirements

Total average flow to biocell (includes return sludge flow and biocell recycle) Biocell pumps rates at 184 L/S	=	12960 m^3/d
Average biocell wetting rate	=	3.4 $L/m^2.S.$

Aeration Basin

Volume	=	230 m^3
Size	=	10 m x 7 m x 3.3 m (SWD)
DT	=	0.8 h
Oxygen requirements	=	8.3 kg/h

Two units of 7.7 kW capacity each are proposed to take care of peak load conditions.

Secondary Settling Tank

Two units each of size 4.5 m x 30.5 m x 3.67 m (SWD) are proposed.

Overflow rate for average flow	=	24.4 $m^3/m^2.d.$
Overflow rate for peak flow	=	61.8 $m^3/m^2.d.$
DT for average flow	=	3.6 h
Solids loading (average flow)	=	85.4 $kg/m^2/m^2.d.$

Chlorine Contact Tank

Volume of chlorine contact tank	=	180 m^3

Size - 10 m x 6 m x 3 m (liquid depth)

Operation of Treatment Plant For Flows Less Than Design

New treatment plants frequently operate at flows less than the design flow rate.

Several options are available in operating a plant at flow rates and organic loadings less than design. These alternatives include: Do nothing, reduce concentration of MLVSS, or take an aeration basin out of service. These alternatives are discussed below using as an example, an ABF plant designed for a system F/M = 1.0 and a MLVSS of 3 000 mg/L, but presently operating at one -half the design influent organic load.

Do Nothing

This is a reasonable alternative. The System F/M would be reduced to 0.50, provided the MLVSS is held at 3 000 mg/L. The plant would achieve nitrification. Aeration basin oxygen requirements would be 50 percent greater than that observed when the plant is achieving only carbonaceous BOD removal. The net sludge production or quantity of sludge to waste would be reduced due to increased destruction of solids when the system nitrifies. Increased foaming would be observed in the aeration basin and more scum would have to be removed from the secondary clarifier.

Reduce Concentration of MLVSS

Reducing the concentration of mixed liquor VSS increases the F/M. Operating the plant at a MLVSS = 1 500 would result in a System F/M = 1.0. The plant would achieve carbonaceous BOD removal. Sludge production would be roughly one-half the estimated design sludge production rate. A minimal increase in effluent suspended solids might be observed due to the leakage of colloidal material, which is normally removed by a higher concentration of MLSS. This occurrence must be confirmed by actually operating the plant at a lower MLSS.

Take an Aeration Basin Out of Service

Some plants have two aeration basins and one of these basins can be removed from service. The net effect of this action would be to increase the F/M to 1.0 and that the plant would essentially be operating under design load conditions.

The final method selected for operating the plant for loads less than design are not dependent upon the characteristics or limitations of the ABF process, but rather other factors specific for each site. Some of these factors include: energy costs, sludge disposal costs and variations in wastewater flow and/or strength characteristics. The best solution is probably the one which provides treatment requirements at lowest total operating costs.

PRELIMINARY RESULTS OF PERFORMANCE

The results of the performance of the treatment plant during August-September, 1984 in respect of the parameters BOD and SS are given on the following page in Table 2.

TABLE 2. TREATMENT PLANT PERFORMANCE DATA
(August-September, 1984)

Date	Primary clarifier influent		Primary clarifier effluent		Final effluent	
	BOD mg/L	SS mg/L	BOD mg/L	SS mg/L	BOD mg/L	SS mg/L
84-08-21[1]	-	123	-	44	-	7
84-08-23[1]	170	159	58	28	7	5
84-08-28[2]	-	147	-	46	-	9
84-08-30[2]	145	117	100	49	20	11
84-09--4[2]	-	145	-	56	-	14

Note: 1) Grab Samples 2) Composite Samples
The treatment plant is functioning well meeting the effluent requirements.

SUMMARY AND CONCLUSIONS

Information on kinetics and relative importance of liquid-phase and slime phase organic removal in the ABF process is not presently available. Thus, process constraints are not completely understood compared to an activated sludge system. However, operating experience with this system has proven that the system is capable of achieving high removals of BOD. The system confers operational stability inherent in a trickling filter system. The use of short-term aeration as part of the ABF process improves process reliability. It has also been found that the heat loss across the system is minimal; a stable heat loss characteristic is advantageous in cold climate operation.

Thus ABF process is an attractive, competitive secondary treatment alternative worth consideration in any evaluation of alternative mechanical treatment systems.

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EPURATION AEROBIE PAR CULTURES FIXEES : PROCEDE BIOFOR

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ABSTRACT

Les techniques de traitement d'eaux utilisant les cultures bactériennes fixées sur différents types de supports sont très étudiées depuis quelques années déjà. Après avoir préalablement comparé différents procédés, notre choix s'est porté sur un réacteur à flux ascendant et en lit fixe : le procédé Biofor. La première partie de cette communication traite des résultats obtenus sur ce type de réacteur alimenté avec une eau résiduaire urbaine ayant subi une décantation primaire.

Les normes de rejet que nous nous étions fixées correspondent à l'actuel niveau e de la législation française :

- DCO = 90 mg/l sur un prélèvement moyen de 24h non décanté
- DBO5 = 30 mg/l sur un prélèvement moyen de 24h non décanté
- MES = 30 mg/l sur un prélèvement moyen de 2h non décanté

Le matériau support sélectionné, lors d'essais préalables, est de la biolite L de taille effective : 2,7 mm.

Nous avons porté notre attention sur l'élimination biologique des pollutions carbonnées et azotées et de l'effet filtrant du réacteur considéré.

Nous avons en outre établi avec précision les paramètres de dimensionnement et optimisé la technologie de mise en oeuvre.

Dans la dernière partie de cette communication sont présentées les deux premières réalisations, Degrémont utilisant ce procédé.

- Royan Saint Palais (France) : 40 000 habitant équivalent (HE)
- Métabief (France) : 11 000 habitant équivalent (HE)

INTRODUCTION

Dans les pays industrialisés, les flux de pollution sont en augmentation constante et les normes de rejet deviennent de plus en plus strictes. En outre, les terrains disponibles pour construire des stations d'épuration sont de plus en plus rares et chers. Aussi les traiteurs d'eau sont-ils amenés à développer des techniques permettant de travailler à des charges volumiques nettement supérieures à celles utilisées en boues activées classiques. Ces dernières techniques connaissent en effet des limitations :

- la concentration en biomasse active
- le transfert d'oxygène vers la bactérie.

En outre l'exploitation des ouvrages de clarification secondaire est délicate car les boues activées des bassins à très forte charge décantent très difficilement.

L'expérience acquise montre qu'il est difficile de dépasser des charges volumiques en DBO5 de 1,5 kg/m3 de bassin/jour.

Aussi le Groupe de la Lyonnaise des Eaux s'est orienté vers un procédé permettant de travailler à des charges nettement plus importantes et d'obtenir une eau traitée de meilleure qualité avec des teneurs en MES et DBO5 inférieures ou égales à 30 mg/l.

Le traitement par cultures fixées semble être la voie la plus intéressante. Il permet de coordonner les deux effets, concentration en biomasse active très importante et activité plus forte d'une biomasse fixée. Le clarificateur secondaire, ouvrage de conduite très délicate sur les stations à boues activées forte charge, est supprimé puisque le réacteur à cultures fixées remplit le double rôle : épuration biologique et filtration.

Avec la mise au point d'un tel procédé, il devient possible de concevoir des stations d'épuration à très forte charge et ce avec un très bon rendement épuratoire.

Suite à l'expérience industrielle acquise sur ce genre de réacteurs (La Tremblade, Briançon et nitrificateur par cultures fixées en flux ascendant) et à de nombreux essais sur pilotes industriels, la mise au point d'un réacteur à biomasse fixée en lit fixe à flux ascendant a été developpé sous l'appellation Biofor.

Le domaine d'application est les eaux résiduaires après décantation primaire.

Les normes de rejets auxquelles doit permettre d'aboutir un tel traitement, correspondent à l'actuel niveau e de la législation française, c'est-à-dire :

- DCO : 90 mg/l sur un prélèvement moyen 24h non décanté
- DBO5 : 30 mg/l sur un prélèvement moyen 24h non décanté
- MES : 30 mg/l sur un prélèvement moyen 2h non décanté

EXPERIMENTATION PILOTE

Matériel et méthode

Le réacteur (Figure 1) utilisé pour les essais est une colonne de 1,05 m de diamètre, d'une hauteur totale de 5,20 m avec une hauteur au-dessus du plancher de 4,20 m. L'air et l'eau, process et lavage sont répartis à la base du réacteur par des buselures, ayant fait l'objet d'une étude particulière pour éviter tout risque de colmatage. L'eau traitée est récupérée au sommet du pilote par surverse dans une goulotte de reprise. Le long de la colonne sont ménagés des piquages permettant à la fois la prise d'échantillons et la mesure des pertes de charge dans le matériau. Il est également possible sur ce pilote de faire varier la hauteur de matériau entre 2 et 3 m.

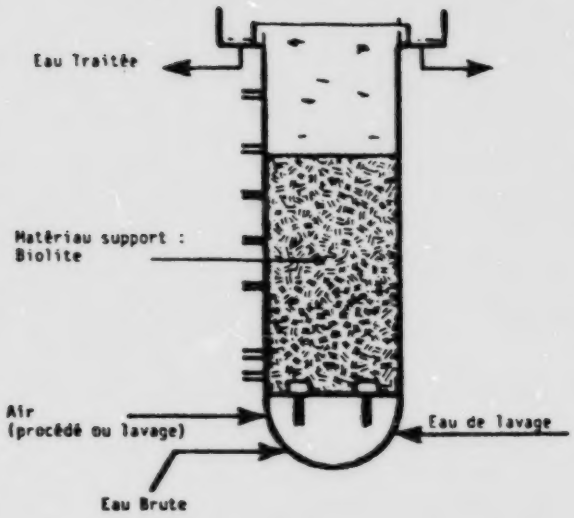


figure 1: SCHEMA DE TRAITEMENT

Le matériau support retenu est de la biolite de type L, de taille effective : 2,7 mm et de coefficient de vide de : 50 %.

Des essais antérieurs ont montré ses capacités tant en filtration qu'en élimination de pollution carbonée en traitement tertiaire (La Tremblade, 1977, Degrémont 1978 (1)), après un traitement physico-chimique (Briançon, 1979, Faup et Jacquart 1981 (2)) et qu'en nitrification par cultures fixées (Elancourt, 1982, Faup et Al, 1982 (3)).

Audic, 1983 (4) Mac Rae, 1983 (5) et Boloorghi, 1983 (6) ont montré qu'il y avait au moins trois critères de choix pour un bon matériau support :

- le volume cumulé de pores et rayon supérieur ou égal à 4 um qui doit êtrele plus grand possible.
- la présence d'hématite qui influencerait de façon importante l'adsorption des bactéries sur un support.
- la rugosité externe qui conditionne la résistance du décrochage de la biomasse.

Audic a montré par ses travaux que la biolite L correspondait d'une façon satisfaisante à ces deux critères et de plus, en nitrification, il a montré que c'était le plus performant parmi les différents matériaux testés (sables, charbons actifs et biolites).

Le réacteur à cultures fixées était installé sur le site de la station d'épuration de Colombes au CERDEG (Centre d'Etudes et de Recherche DEGREMONT) et nous avons utilisé l'eau décantée primaire fournie par cette dernière.

La qualité de l'eau brute durant les essais est résumée dans le tableau ci-après.

TABLEAU 1 : QUALITE DE L'EAU BRUTE

	moyenne	écart type	nombre de valeur
DCO totale (mg/l)	200,8	74,8	241
DCO soluble (mg/l)	99,9	45,9	324
MES (mg/l)	81,9	25,7	422
DBO_5 (mg/l)	99,8	46,1	63

Les débits d'eau testés à l'entrée du réacteur ont été : 3 puis 4,5 et 6 m3/m2/h. Les débits d'air associés ont varié de 6 Nm3/m2 de secteur/h jusqu'à 12 Nm3/m2 de section/h avec des rapports air/eau égaux selon les cas à 1 - 1.5 et 2 volumes d'air/volume d'eau. Les charges volumiques appliquées en DCO totale, correspondant aux charges hydrauliques étudiées, ont été de 4 à 15 kg/m3 de matériau/jour.

- Mesure de l'activité respiratoire

Les mesures de consommation d'oxygène sont réalisées à l'aide d'un respiromètre de type Warburg. Chaque résultat est établi à partir des variations de pression mesurées sur 3 fioles témoins et 7 fioles contenant les fractions représentatives de l'échantillon, soit environ 1 gramme de support colonisé (Degrémont (1)).

- Autres méthodes de mesure

Les MES, les DCO et les DBO5 sont déterminées avec des méthodes normées par l'AFNOR sous les références suivantes :

. les MES norme AFNOR T90/ SC 1
. les DCO norme AFNOR NFT90/101
. les DBO_5 norme AFNOR NFT9/103

Résultats obtenus et discussion

a) Mesure d'oxygène dissous dans le réacteur. Des profils de concentration d'oxygène dissous ont été effectués sur toute la hauteur du filtre. Ces mesures ont été réalisées avec un réacteur ensemencé et nous n'avons pas noté de variations importantes dans les profils entre les débuts et fin de cycle. La figure 2 résume les valeurs mesurées. Le profil moyen a été estimé à partir de la moyenne des concentrations d'oxygène à une même hauteur pour des vitesses d'eaux variant de 3 à 6 m3/m2/h et un rapport air/eau variant de 1 à 2. De part et d'autre de ce profil sont indiqués les 2 profils extrêmes afin de mieux mettre en évidence les variations de concentration d'oxygène à une hauteur donnée. Ces valeurs semblent indiquer que l'activité de la biomasse résidente dans le réacteur n'est pas limitée par la concentration en oxygène dissous.

①PROFIL A :
VITESSE D'EAU : 3m³/m²/h
VITESSE D'AIR : 3 Nm³/m²/h

②PROFIL MOYEN

③PROFIL A :
VITESSE D'EAU : 6 m³/m²/h
VITESSE D'AIR : 12 Nm³/m²/h

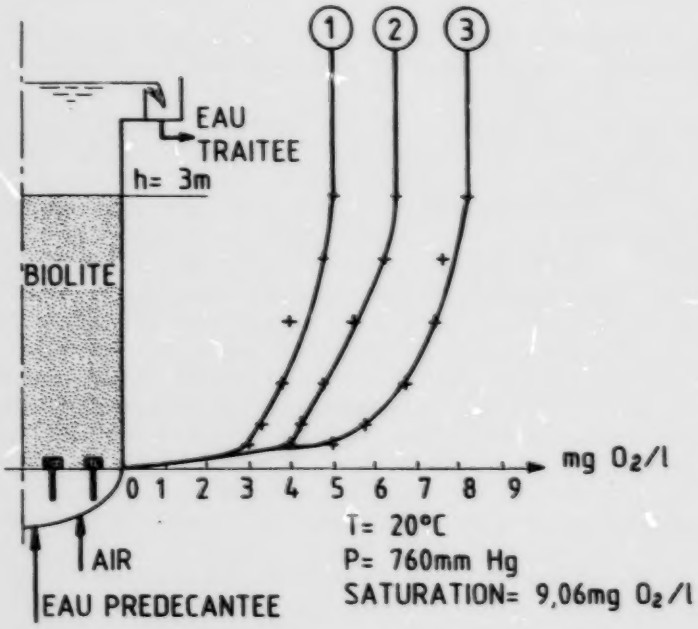


figure 2:
PROFIL DE CONCENTRATION D'OXYGENE DISSOUS
SUR TOUTE LA HAUTEUR DU REACTEUR
VITESSE D'EAU: 3 à 6 m³/m²/h
RAPPORT AIR/EAU: 1 á 2

TABLEAU 2 : EVOLUTION DE L'ELIMINATION DES MES EN FONCTION DES HAUTEURS DE BIOLITES ET DES VITESSES D'EAU ETUDIEES

Hauteur biolite (m)	Vitesse d'eau (m3/m2/h	Teneur en MES eau décantée (mg/l)	Teneur en MES eau traitée (mg/l)	Rendement (%)
2	3	78	12	85
	4,5	74	15	80
	6	69	16	77
3	3	83	18	78
	6	92	21	77

Cette première hypothèse est confirmée par les mesures d'élimination de la DCO soluble in situ qui ne montrent pas de limitation due à l'oxygène.

b) Action sur la pollution particulaire. Un des objectifs de cette étude était d'obtenir une teneur maximum en MES dans l'eau traitée de 30 mg/l. Nous avons suivi l'évolution de ce paramètre pour des vitesses d'eau de 3 puis 4,5 et 6 m3/m2/h avec une hauteur de biolite de 2 m puis 3 m. Le tableau 2 résume les différentes valeurs obtenues.

c) Action sur la pollution soluble. Nous disposons de l'évolution de 3 paramètres pour caractériser cette épuration. La DCO totale et la DBO5 totale sont des paramètres qui englobent les pollutions solubles et particulaires. Pour ces raisons, nous avons préféré suivre l'évolution de la DCO soluble.

La figure 3 donne l'évolution du rendement moyen d'élimination de la DCO soluble (%) en fonction des vitesses d'eau étuidiées. Nous mettons en évidence une baisse du rendement d'élimination de 70 à 50 % quand la vitesse d'eau passe de 3 à 6 m3/m2/h. Cette diminution est imputable à une limitation de l'activité biologique par une baisse de la teneur en DCO soluble dans l'eau pré-décantée (période estivale).

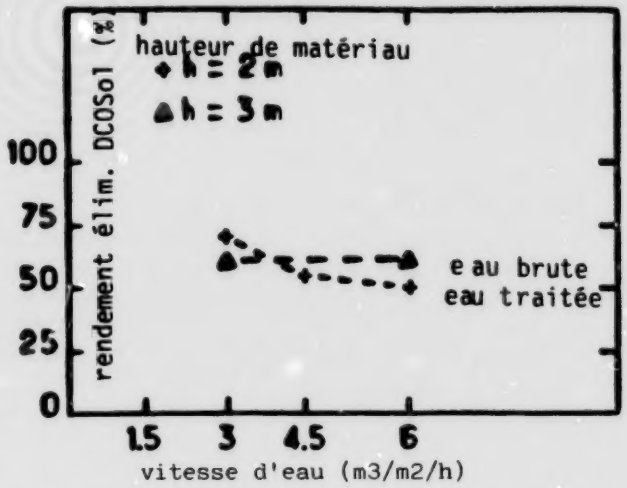


figure 3: Evolution du rendement d'élimination en DCO Sol en fonction des vitesses d'eau étudiées.

Afin de mettre en évidence l'activité microbiologique au sein de la biolite, nous avons dosé la teneur en DCO soluble de l'eau à différentes hauteurs dans la biolite. Les profils de DCO soluble obtenus sont linéaires, ce qui semble signifier que l'activité globale est uniformément répartie sur toute la hauteur du matériau (figure 4). La figure 5 donne la quantité de DCO soluble éliminée/m de matériau en fonction de la concentration en DCO soluble de l'eau brute. Nous pouvons voir qu'il y a une relation linéraire entre ces deux paramètres, par contre, la vitesse de filtration n'a pas d'influence.

Pour confirmer qu'il n'y a pas de gradient d'activité de biomasse fixée sur toute la hauteur du réacteur, nous avons prélevé des échantillons de biolite à différentes hauteurs dans le filtre et avons mesuré leur activité respirométrique dans un appareil de type Warburg. Nous obtenons une valeur moyenne de $8,36 \times 10^{-2}$ ml 02 consommé/h/g de biolite avec un écart type de $0,12 \times 10^{-2}$. Ces mesures indiquent que l'activité de la biomasse fixée est la même sur toute la hauteur du matériau.

De plus les taux de respiration endogène mesurés sur nos cultures fixées ont une valeur de 0,29 g oxygène consommé/j/g/MVS contre 0,13 g d'oxygène consommé/j/g de MVS habituellement obtenus sur une culture libre travaillant à charge massique égale, ce qui confirme bien que la flore bactérienne travaille en forte charge.

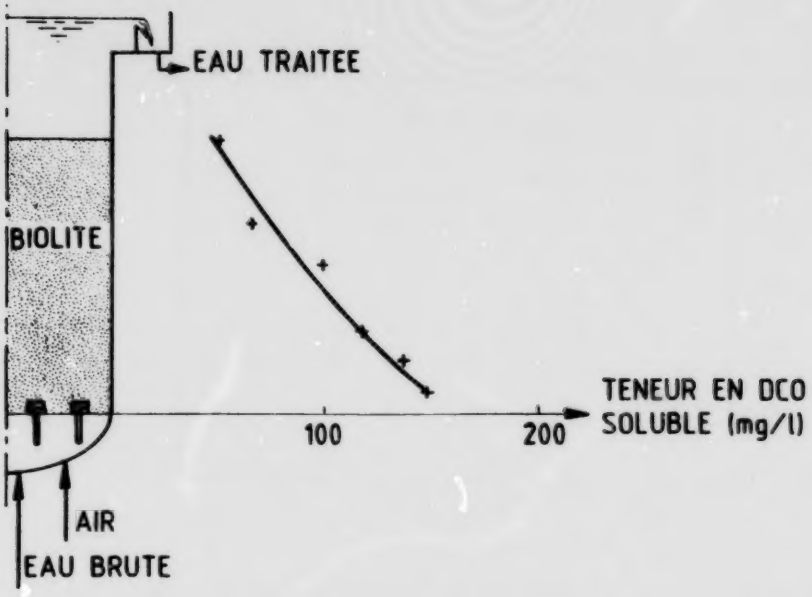


figure 4 :
PROFIL DE DCO SOLUBLE
VITESSE D'EAU : $6m^3/m^2/h$
VITESSE D'AIR : $12\ Nm^3/m^2/h$
HAUTEUR DE BIOLITE : 3m

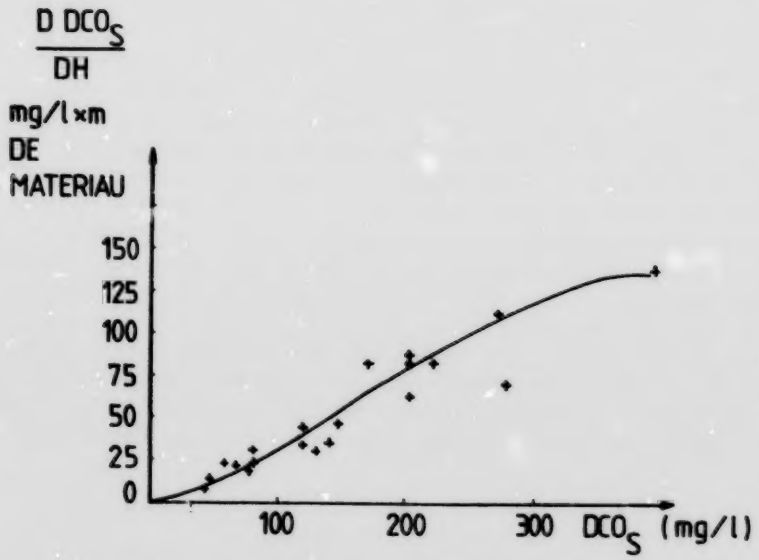


figure 5 : RELATION ENTRE LE TAUX D'ELIMINATION DE DCO_S ET LA TENEUR DE L'EAU EN DCO_S

d) Charges appliquées par cycles en MES, DCO totale et DBO5. Les figures 6, 7 et 8 montrent les variations de charges appliquées en MES, DCO totale, DCO soluble et DBO5 kg/m3 de matériau/cycle, en fonction des vitesses d'eau de 3, puis 5 et 6 m3/m2/h avec 2 m de matériau.

Les charges appliquées par cycle sont une moyenne des perforances du réacteur sur au moins 15 cycles consécutifs. Nous voyons une évolution similaire des charges appliquées en fonction des vitesses d'eau qu'il s'agisse de pollution insoluble ou soluble.

La durée d'un cycle est définie comme le temps de filtration pendant lequel l'eau traitée est conforme aux normes de rejet fixées lors de notre étude.

Ces durées avec 2 m de matériau pour des vitesses d'eau de 3 puis de 4,5 et 6 m3/m2/h sont d'une moyenne de 50 heures pour les 2 vitesses les plus basses et de 27 heures pour la vitesse la plus haute. Il semble donc qu'avec une vitesse de procédé de 4,5 m3/m2/h, la capacité de rétention de la pollution soit maximale.

Avec 3 m de biolite les charges appliquées en kg MES, en kg DCO et en kg DBO5/m3 de matériau/cycle sont restées inchangées. Par contre les durées de cycle ont augmenté de 50 % pour une qualité d'eau traitée similaire. Ces paramètres nous on permis de concevoir à l'échelle industrielle les 2 premières installations de ce type.

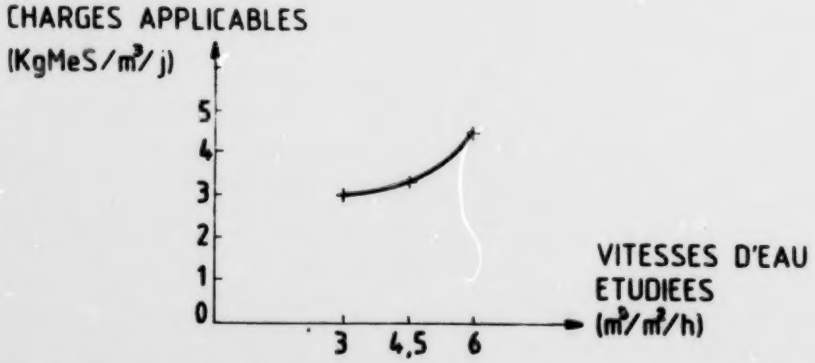


figure 6: EVOLUTION DES CHARGES APPLICABLES EN KgMeS/m³/j EN FONCTION DES VITESSES D'EAU ETUDIEES ET AVEC UNE HAUTEUR DE BIOLITE EGALE A 2m.

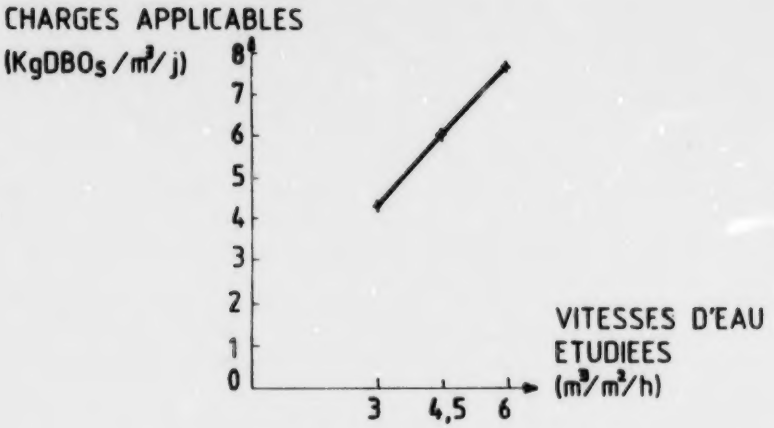


figure 7: EVOLUTION DES CHARGES APPLICABLES EN Kg DBO_5 /m^3/ j EN FONTION DES VITESSES D'EAU ETUDIEES($m^3/m^2/h$) AVEC UNE HAUTEUR DE BIOLITE EGALE A 2m.

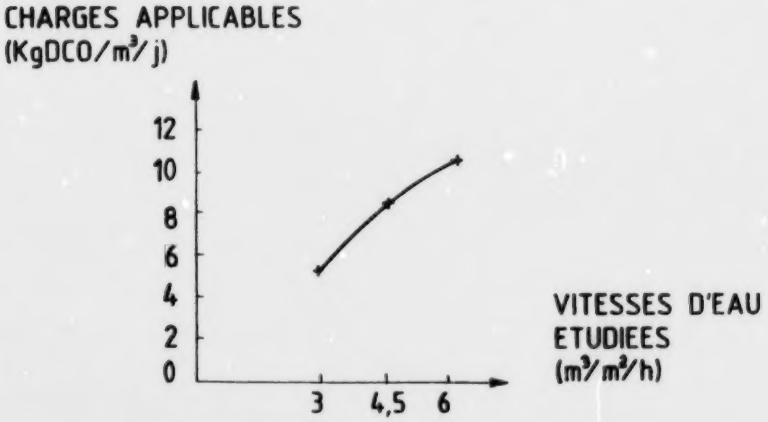


figure 8:
EVOLUTION DES CHARGES APPLICABLES EN KgDCO TOTALE/m^3/ j EN FONCTION DES VITESSES D'EAU ETUDIEES ($m^3/m^2/h$) ET AVEC UNE HAUTEUR DE BIOLITE EGALE A 2m.

e) Evolution des pertes de charge et lavages. Nous appelons perte de charge totale en fin de cycle, la perte de charge dans le réacteur en fin de cycle, c'est-à-dire au moment où il y a dégradation de la qualité de l'effluent traité par relargage de matières en suspension. Avec 2 m de biolite la perte de charge totale fin de cycle est égale à 80 cm de colonne d'eau et avec 3 m de hauteur de biolite nous avons 120 cm de colonne d'eau.

Le lavage s'effectue à co-courant air-eau. Son principe est dérivé des lavages des filtres pour le traitement d'eau potable. Différentes

expérimentations nous ont permis de retenir les conditions de lavage suivantes :

- Une première séquence de brassage avec un débit d'air de 55 à 70 Nm3/m2/h et un débit d'eau faible 8 m3/m2/h.
- Une deuxième séquence d'élution des boues retenues dans la biolite avec un débit d'eau de 25 m3/m2/h et un faible débit d'air 10 Nm3 afin de faciliter cette élution. Les durées des séquences sont de 20 minutes pour la première et 10 minutes pour la deuxième. Le lavage s'effectue avec notre eau traitée et son volume n'excède pas 5 % du volume d'eau filtrée par cycle.

Après le lavage, il reste une population bactérienne fixée suffisante pour éviter une phase de maturation au redémarrage du réacteur. Le rendement d'élimination en DCO soluble est alors égal à celui observé en cours de cycle.

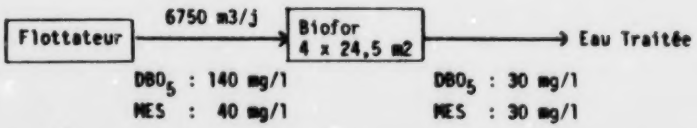
Des bilans de boues ont été effectués sur les eaux de lavages à la fin des cycles. L'étude de 100 bilans montre que la quantité de boues en excès produite est égale à 1 kg de boue/kg DB05 éliminée.

Plusieurs déterminations du taux de matières volatiles dans les boues ont donné le résultat suivant : 80 % de matières volatiles sèches (VSS). C'est un taux courant sur les boues biologiques des procédés boues activées fortes charges.

REALISATIONS INDUSTRIELLES

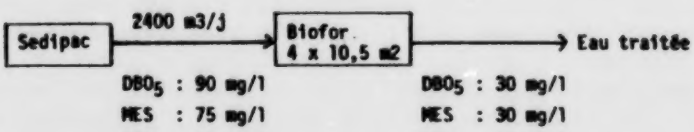
Cette étude a permis outre la détermination des paramètres de dimensionnement, la mise au point de la technologie de mise en oeuvre de ce procédé. Les 2 premières réalisations Degrèmont de ce type sont mises actuellement en service. Nous donnons ci-dessous les schémas de traitement correspondant à chacune de ces stations.

Royan St Palais (45 000 HE)



Charges appliquées : DBO_5 3,2 kg/m3/j
MES 1 kg/m3/j

Metabief (16 000 HE)



Charges appliquées : DBO_5 1,6 kg/m3/j
MES 1,1 kg/m3/j
NTK 0,75 kg/m3/j

Exemple de St Palais

La station de St Palais est située au centre de la presqu'ile d'Arvert, à proximité de la ville de Royan sur la côte atlantique française.

Elle comporte actuellement trois tranches de traitement des eaux usées, pour un total de 150.000 ha.equ.

La dernière tranche, d'une équivalence de 40.000 he doit répondre à la surcharge estivale due à l'afflux touristique. Le fonctionnement est donc de 3 mois par an. Par ailleurs, pour maintenir dans de bonnes conditions de fonctionnement les 2 premières tranches, conçues selon les procédés classiques, il est prévu de leur éviter autant que possible les surcharges quotidiennes de débits. Celles-ci iront donc pour la plus grande partie vers la 3ème tranche.

La qualité de rejet exigée porte sur les matières en suspension : moins de 30 mg/l et sur la DB05 : moins de 30 mg/l.

Compte tenu de ces données, la ligne de traitement suivante a été choisie :

1. Prétraitement (dégrillage-dessablage)
2. Flottation
3. Epuration par cultures fixées aérobies (Procédé Biofor)
4. Désinfection et rejet en mer.

Pour l'étape 3, parmi les techniques à cultures fixées envisageables : préoxygénation, filtration à sec notamment, c'est le procédé Biofor qui convenait le mieux. En effet, si l'eau en sortie de flottateur contient relativement peu de MES (40 à 60 mg/l) elle est par contre très chargée en DB05 (= 140 mg/l).

Il fallait donc une aération efficace doublée d'un temps de contact suffisant.

Dimensionnement du Biofor

La vitesse moyenne de passage de l'eau est 3 m/h avec un coefficient de pointe prévue de 2,75, ce qui donne une vitesse maximale de 8,2 m/h.

Quatre réacteurs rectangulaires se partagent une surface totale de filtration de 98 m2.

La hauteur de garnissage, calculée par des équations obtenues lors des essais pilote est de 2,7 m de biolite L.

Premiers résultats de la station

La mise en route a eu lieu au début de Juillet 1984.

Dans une première période le matériau est ensemencé pour atteindre en 4 semaines des activités biologiques et physico-chimiques stables et satisfaisantes. Cette maturation, c'est important, n'est pas à refaire chaque année mais seulement lorsque le matériau est vierge. Les essais pilotes ont toujours confirmé que avec un matériau tel que la biolite L, l'activité repartait en quelques heures après un arrêt,si longue que soit la durée.

Au-delà de cette première période, les caractéristiques moyennes obtenues sur l'eau traitée sont les suivantes :

MES	10 - 17 mg/l
DCO	30 - 50 mg/l
DB05	<10 - 25 mg/l

Les exigences du traitement sont donc très largement respectées. De plus des essais ont été réalisés à différentes vitesses en eau : jusque 8,4 m/h.

Il a été constaté, comme lors des essais pilote que l'activité biologique ne s'en trouvait pas perturbée. Cependant la teneur en MES de l'eau traitée augmente jusque 20 - 28 mg/l.

En ce qui concerne les caractéristiques de conduite de l'installation, les principales valeurs obtenues lors des essais pilotes ont été réobtenues.

Les durées de cycle variant de 24 à 48 heures, voire plus, selon la qualité de l'eau.

En fin de cycle la perte de charge est de 1,1 - 1,2 m C.E.

Les charges appliquées sont de 4 à 5 kg/m3.j en DBO5 (lorsque la concentration de l'eau le permet) et de 1,5 à 2 kg/m3.j en MES.

La consommation d'eau par lavage est d'environ 8 m3/m2 de réacteur.

En conclusion les premiers mois de fonctionnement de cette station sont encourageants, les normes de rejets sont largement respectées et les résultats obtenus concordent pleinement avec ceux des essais réalisés sur pilote.

CONCLUSION

Cette étude met en valeur les performances du procédé par cultures fixées pour traiter une eau résiduaire décantée primaire. Les charges appliquées et la qualité de l'eau traitée sont supérieures à celles observées sur des traitements à boues activées forte charge. Ces performances supérieures, observées en cultures fixées, peuvent être une conséquence à l'augmentation de biomasse active dans le réacteur et/ou une activité de la biomasse fixée plus forte que celle de la biomasse libre. L'exploitation de ces stations permettra de mieux cerner l'intérêt industriel de ce procédé qui devrait dans l'avenir occuper un niveau important dans l'arsenal des techniques disponibles.

Des études comparatives ont été menées sur ce sujet. Audic (1983) a comparé les activités spécifiques de Nitrobacter en culture libre et fixée. Il a mis en évidence une activité spécifique supérieure de Nitrobacter fixé sur un support.

Actuellement à l'université de Metz, des recherches ont lieu afin de comprendre les mécanismes d'élimination de la pollution carbonée en lit fixe. Dans ce cadre, la caractérisation du biofilm, la mise en évidence de nouveaux critères pour un support et l'optimisation du procédé seront étudiées.

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THE USE OF ARTIFICIAL CATTAIL MARSHES TO TREAT MUNICIPAL WASTEWATER IN A NORTHERN ONTARIO COMMUNITY

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ABSTRACT

Many small northern communities cannot sustain operating costs of conventional secondary sewage treatment facilities. In addition they may have poorly maintained combined sewer systems which result in dilute effluent and thus drive the capital cost of conventional treatment to prohibitive levels.

In a jointly funded project, the Ontario Ministries of Environment and Northern Affairs have been studying the potential for use of artificial cattail marshes in such applications.

A 0.1 ha test plot was established at Cobalt, Ontario and loaded with dilute raw sewage for a two year period. It was hand planted with cattails and operated on a continuous discharge basis in all seasons. Results show that the marsh system is capable of producing an effluent which meets the provincial secondary sewage treatment objectives.

INTRODUCTION

In July of 1980 the Ontario Ministry of the Environment (M.O.E.) started the Listowel Marsh Project in southwestern Ontario (Figure 1). The purpose of the project was to experimentally investigate the potential for using artificial cattail marshes to treat municipal sewage. Although the work at Listowel is still on-going, interim assessments have established that properly configured cattail marshes have the capacity to significantly improve the water quality of sewage wastewaters. (Black, et al. 1981).

In 1981 inquiries were made to M.O.E. from the Ontario Ministry of Northern Affairs as to the feasibility of marsh treatment technology for communities in Northern Ontario. Many small northern communities cannot sustain the operating costs of conventional secondary sewage treatment facilities. In addition they may have poorly maintained combined sewer systems which result in dilute effluent and thus drive the capital cost of conventional treatment to prohibitive levels. The northern community of Cobalt (Figure 1) was suggested as one location which had such sewage treatment problems.

The Cobalt Marsh Project was created that same year with the objective of determining if the marsh treatment technology could be applied in more extreme northern climates. The project was jointly supported by the Ontario Ministries of Northern Affairs and Environment.

FIGURE 1: LOCATION MAP FOR THE LISTOWEL AND COBALT ARTIFICIAL CATTAIL MARSHES. THE ISOLINES SHOW THE LENGTH OF THE GROWING SEASON IN DAYS.

METHOD

A test plot was constructed based on the design of the best performing systems at the Listowel facility. The plot was built out of insitu materials and lined with clay from off-site. The plot was approximately 310 m in length and 3 m wide giving a length/width ratio of greater than 100:1. It had a serpentine configuration.

The marsh was established in July of 1982 by planting individual cattail shoots with attached pieces of rhizome at one meter intervals throughout each plot. The cattails were removed from adjacent natural stands.

The flow of sewage was controlled by a flow control structure located at the sewer outfall near the community's arena. The bulk of the sewage by-passed the plot by over-flowing a large rectangular weir which acted to dampen the severe fluctuations in head which occurred in the sewer system. The flow of sewage to the plot was controlled by a 22½ degree V-notch weir in the same structure. The head of water behind the weir was measured by a float device and Stevens stage recorder. The calibration of stage records was maintained by the taking of timed bucket discharges during regular visits to the site. An identical weir and stage recorder were installed in the flow control structure at the discharge from the plot.

The design called for the plot to be loaded at approximately 0.2L/sec. (17 m^3/day, 20.0 mm/day) which gave an areal loading similar to that used at Listowel. Depth was to be maintained at approximately 15 cm all year round to keep the retention time near the 7 day period used at Listowel. Depth was increased in winter time to allow 15 cm of water to be maintained under an ice layer.

The raw sewage influent to the marsh system and the effluent from the marsh plot (if present) were sampled weekly from July 11, 1983 until the 28th of November and then alternate weeks through the winter months. The samples were analysed for BOD_5, suspended solids, filtered total nitrates, filtered total ammonia, total Kjeldahl nitrogen, pH, conductivity at 25 degrees C and sulphates.

The bacteriological quality of the water was sampled and counts were taken of faecal streptococci, fecal coliforms and total coliforms. From July 11 to August 8, 1983 only the effluent from the plot was analysed. Subsequently, samples of both influent and effluent were monitored.

In late November the Ministry of Health Lab changed its dilution technique and method of reporting. As a result the maximum level of detection for bacterial counts was raised an order of magnitude but total coliform counts were no longer done.

The bacteriological samples were shipped on the date of sampling packed in ice packs but not always submitted to the lab the next day. Due to variation in ambient temperature and handling, samples taken on different dates may not be directly comparible.

RESULTS

Biology of the Marsh

The growth and development of the cattails was assessed on July 5, 1983 approximately one year after their planting. As in Listowel the marsh had filled in quickly and there was evidence of both seedling establishment and continued vegetative reproduction. The cattails were approximately 1.3 m high and while there were some small voids the average density was estimated to be 20 stems/m^2.

Parallel to the Listowel experiment the marsh plot had already been invaded by several other species. Unidentified species of bullrush (Scirpus) and smartweed (Polygonum) were well established.

During the winter of 1982-83 muskrats tunneled through the berms of the last channel of the plot. As a result the channel dried out in the early spring and the cattails died. The holes were sealed in May and the area reflooded. Some replanting was done but it proved superfluous as new seedlings had vigorously recolonized the area by the July 5, 1983 inspection.

Hydrology of the Marsh

Attempts to measure the incoming and outgoing flows of the experimental plot were fraught with difficulties. The stage of the water in the splitter box, behind the V-notch weir, was subject to great fluctuations despite the dampening effect of the rectangular weir. As well, solids accumulated in the splitter box which interfered with both the floats on the stage recorder and the V-notch weir itself. There were also failures in the stage stage recorder mechanism itself which resulted in lost data.

The recorder charts that were recovered have not been digitized at the time of this report. However, there are numerous bucket discharge measurements available which can be used to characterize flows.

These data indicate that the marsh was loaded at a much higher rate than was originally intended. Initially the marsh received approximately three times the design volume but through time this gradually increased to five times the loading.

The outflow of the marsh always was found to be substantially less than the inflow. This was especially true during periods of the fall where stop logs had been put in place to increase the water level.

Water Quality of Influent and Effluent

The results of the water quality sampling are presented in Tables 1 and 2. $B.O.D._5$, as also presented graphically in Figure 2.

The BOD_5 in the raw sewage averaged 21.7 mg/L (Table 1) Passage through the marsh reduced the BOD_5 by an average of 80.0%, producing an effluent with an average 4.1 mg/L BOD_5. The final product always contained less than 15 mg/L BOD_5, the objective of secondary sewage treatment.

TABLE 1: RESULT OF WATER CHEMISTRY SAMPLING FOR BOD_5 SS, TP, AND CHLORIDE.

DATE	BIOLOGICAL OXYGEN DEMAND mg/L		SUSPENDED SOLIDS mg/L		TOTAL PHOSPHOROUS mg/L		CHLORIDE mg/L	
	on	off	on	off	on	off	on	off
1983								
July 11	16.6	<2.0	19.7	3.6	1.40	1.23	24.0	23.4
18	27.0	1.7	42.5	4.5	1.76	0.40	96.6	23.6
25	13.2	2.0	19.5	0.1	1.35	0.36	45.4	48.6
Aug. 2	33.0	3.4	38.6	5.8	1.85	0.44	38.6	25.6
8	17.6	4.7	26.1	5.1	-	-	25.6	24.0
15	35.0	11.0	30.4	2.8	2.21	0.49	21.4	18.6
22	18.4	6.1	51.2	3.2	1.93	0.46	19.8	18.0
29	9.3	1.4	20.5	2.7	1.70	0.23	18.4	15.0
Sept. 6	17.0	2.3	21.4	4.7	0.78	0.20	16.2	16.2
14	33.0	5.4	24.9	105.0	1.90	0.27	17.6	0.2*
20	25.0	1.9	49.5	2.7	0.72	0.36	14.4	18.4
26	27.0	1.4	52.2	2.3	1.25	0.44	22.4	22.6
Oct. 4	6.4	2.5	11.5	8.2	0.60	0.25	32.2	19.6
11	27.0	2.9	37.5	7.9	1.00	0.43	36.0	27.4
18	16.4	1.3	16.6	7.4	1.70	0.70	22.8	24.6
24	21.0	1.4	17.8	23.7	2.25	0.70	18.8	20.4
31	26.0	2.2	32.3	24.2	1.80	0.80	16.8	17.8
Nov. 7	48.0	4.7	38.9	21.0	2.78	1.03	29.2	20.6
14	63.0	6.3	55.4	-	5.00	1.45	34.4	53.6
21	16.7	8.0	61.2	59.9	1.30	1.25	122.6	19.6
28	36.0	7.0	26.1	75.8	3.05	1.00	31.2	47.8
Dec. 12	27.0	11.3	16.5	270.0	1.40	1.30	17.8	22.6
1984								
Jan. 3	23.0	9.4	13.5	16.1	0.95	0.95	19.8	15.2
16	10.1	6.1	19.7	7.2	1.15	0.68	167.2	167.4
Feb. 13	23.4	3.1	281	7.8	0.80	0.70	288	61.6
Mar. 13	27.0	-	23.8	-	0.77	-	19.8	-
27	27.0	10.8	34.1	20.1	1.10	0.91	29.4	30.6
Apr. 9	14.2	10.7	14.9	123.0	-	1.88	62.0	78.4
23	3.3	3.2	4.7	3.6	0.22	0.17	44.0	50.0
May 7	21.3	3.1	24.1	4.1	2.00	0.33	31.8	33.6
15	12.1	0.6	9.5	3.6	9.5	3.6	30.2	31.8
28	17.0	3.2	21.7	12.0	1.20	0.18	40.2	48.4
June 5	13.7	3.1	17.1	5.3	1.54	0.62	30.6	33.2
12	22.0	1.8	23.5	5.3	1.77	0.36	33.0	27.4
19	6.1	2.5	8.2	2.7	0.76	0.38	29.0	21.6
25	11.3	2.1	19.1	4.7	0.62	-	36.8	24.2
Jul. 3	35.6	2.5	9.6	3.0	1.28	0.54	26.0	25.4
9	19.0	1.6	13.9	3.4	1.50	0.35	22.6	19.0
16	11.9	3.4	14.3	7.8	1.40	1.03	22.0	20.2
24	8.6	3.6	10.3	7.9	1.70	0.76	36.8	24.2

-- No data available
*- anomalous data point - inconsistent with conductivity data and other ambient chloride concentrations

TABLE 2: RESULTS OF BACTERIOLOGICAL SAMPLING PROGRAM

DATE	FAECAL STREPTOCOCCI x 10/L		FAECAL COLIFORMS x 10/L		TOTAL COLIFORMS x 10/L	
	on	off	on	off	on	off
1983						
July 11	-	9,000	-	12,000	-	-
18	-	8,000	-	280	-	-
26	-	9,000	-	+++	-	+++
Aug. 2	-	3,840	-	1,200	-	6,400
8	-	<40	-	40	-	2,800
15	+++	<100	+++	+++	+++	+++
23	-	- -	+++	40	+++	400
29	+++	4,000	+++	960	+++	6,000
Sept. 6	+++	+++	+++	80	+++	1,000
14	+++	+++	1,800	+++	3,000	+++
20	+++	3,080	+++	40	+++	4,800
26	+++	1,160	+++	40	+++	3,200
Oct. 4	3,600	<40	+++	600	+++	3,000
11	+++	480	+++	40	+++	600
18	+++	400	+++	40	+++	200
24	+++	<40	+++	40	+++	200
31	+++	<200	+++	320	+++	400
Nov. 7	2,680	1,720	+++	2,560	+++	6,400
14	+++	2,000	+++	3,400	+++	20,000
21	+++	+++	+++	+++	+++	+++
28*	>200,000	6,600	>600,000	1,000	-	-
Dec. 12	60,000	60,000	60,000	14,000	-	-
1984						
Jan. 3	>200,000	62,000	290,000	10,000	-	-
16	8,000	2,800	86,000	16,000	-	-
30	16,000	5,400	44,000	20,000	-	-
Feb. 27	31,000	2,000	27,000	2,200	-	-
Mar. 13	22,000	-	220,000	-	-	-
27	1,800	3,000	53,000	20,000	-	-
Apr. 9	18,000	4,000	150,000	<10,000	-	-
23	15,000	800	>200,000	3,000	-	-
May 7	200,000	24,000	>600,000	<10,000	-	-
15	34,000	<200	250,000	<10,000	-	-
22	130,000	1,000	450,000	10,000	-	-
29	>200,000	800	>600,000	<10,000	-	-
June 12	130,000	3,200	>600,000	<10,000	-	-
19	70,000	24,000	360,000	<10,000	-	-
25	40,000	200	11,000	<10,000	-	-
July 3	200	110,000	340,000	<10,000	-	-
9	30,000	10,000	410,000	<10,000	-	-
16	50,000	800	520,000	<10,000	-	-
24	110,000	1,600	600,00	<10,000	-	-

+++ - Denotes greater than maximum detectable number
Faecal strep. 20,000
Faecal coliform 12,000
Total coliform 80,000

* - Samples heated in transit

\- - No data available

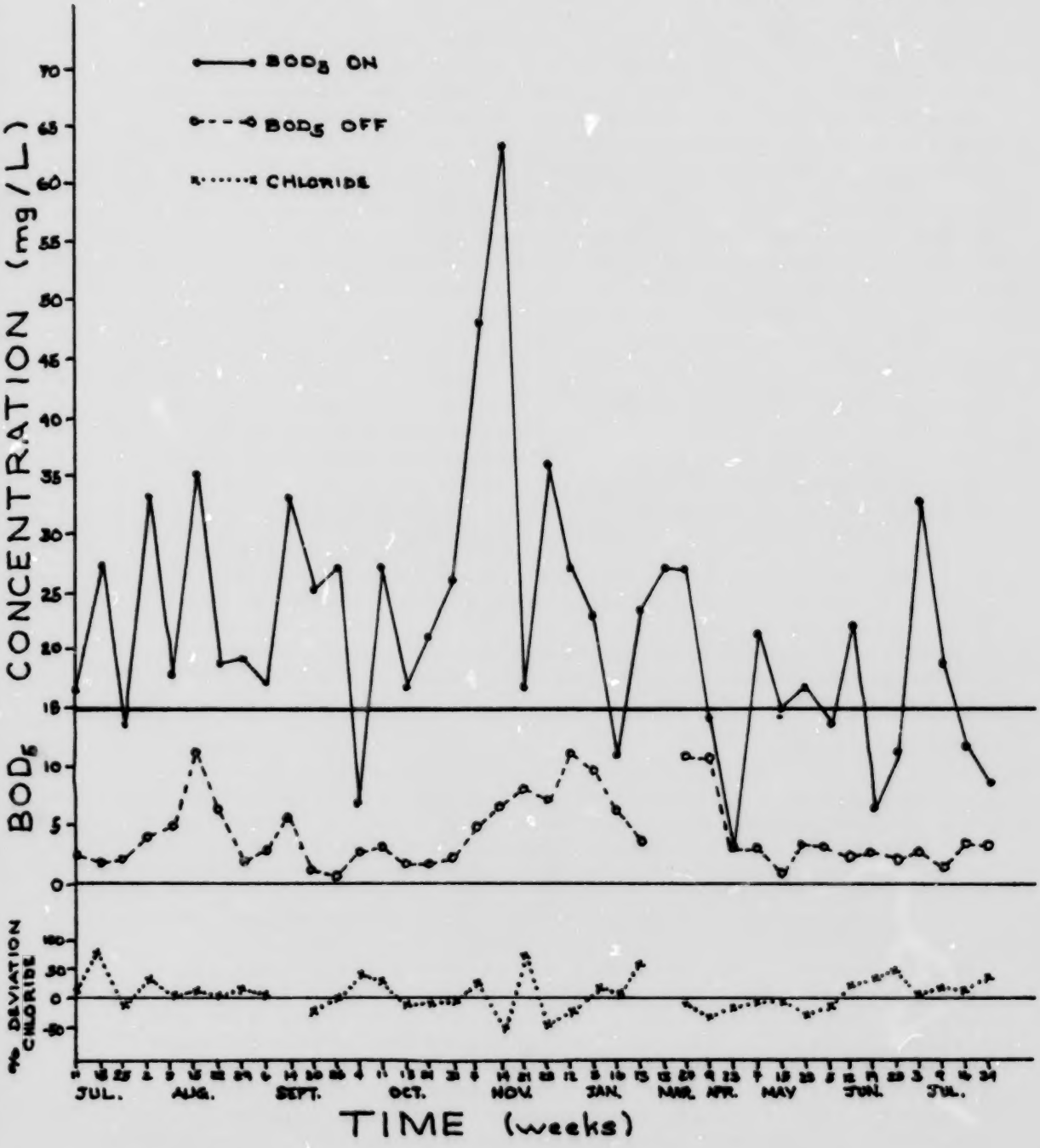

FIGURE 2: A GRAPH OF BOD_5 CONCENTRATIONS IN THE INFLUENT AND EFFLUENT OF THE MARSH

Suspended solids in the raw sewage averaged 31.9 mg/L (Table 1). Excluding three anomalies in the data the average SS concentration in the effluent was 10.9 mg/L. The objective for SS levels in a secondary effluent is 15 mg/L.

The total phosphorous in untreated sewage averaged 1.74 mg/L (Table 1). The outflow from the marsh contained an average 0.74 mg/L TP, which is an average 55.4% less than in the raw sewage. The objective for phosphorous levels in a secondary effluent in the Great Lakes Basin is 1 mg/L.

Table 2 reports the results of the bacteria counts. On September 14, 1983 the data reported was anomalous. The values reported were the exact reverse of the values reported in both the weeks before and after. Since it is a reasonable possibility that the samples were accidentally switched on this date, these data are not considered in the analysis.

The results show that while bacterial counts are consistently high going into the plots the output counts vary. Out of 33 paired input and output samples counted for faecal streptococci, 29 showed lower bacterial populations in the outflow. Three could not be distinguished, since both input and output levels were above the maximum detectable limit of the dilution series. One sample showed a slight increase in the faecal streptococci count. Similarly, faecal coliforms were lower in the outflows in 32 out of 33 paired comparisons with one pair being undistinguishable.

The numerical value of the counts show that the numbers of faecal streptococci and faecal coliform discharging were usually low during cool periods. In the best of conditions the marsh is capable of discharging an effluent which meets swimming and bathing water quality objectives. (Ministry of the Environment, 1978).

DISCUSSION

The biological growth and development of the marsh parallelled the marsh development at Listowel. The height and density of the cattails after one year were not as great as in Listowel but the vigor with which the marsh established itself was the same.

The hydraulic integrity of the marsh however was not up to the standards of the Listowel facility. The bucket discharge data indicates that there was a large loss of fluid from the plot due to exfiltration despite its clay lining. Fortunately this loss does not interfere with the critical criteria for the assessment of the performance of the marsh since these are expressed as concentrations and not loadings. Exfiltration cannot decrease the effluent concentration, but, it may increase it in the case of suspended solids and bacteria.

One consequence of the presence of an unknown level of exfiltration, however, is that one cannot be certain to what extent the improvement in the contaminant concentration seen through the marsh is a function of dilution (due to rainfall or snow melt) or concentration (due to evapotranspiration losses). For this reason the deviation in chloride concentrations were used to calculate an index of the potential

dilution/concentration by the following formula:

$$\text{Percent Deviation} = \frac{\text{Input Cl} - \text{Output Cl}}{\text{Input Cl}}$$

Thus, high positive values indicate an apparent period of dilution and high negative values a period of concentration. The index is presented on Figure 2 as an aid in interpretation of the data presented. It can be seen from this that during most of the study period dilution and concentration were not related to effluent improvement.

As illustrated by Figure 2 the marsh maintained an effluent BOD_5 concentration well within the 15 mg/L standard expected of a secondary sewage treatment system during the whole study period. As was found in Listowel the effluent quality of the marsh seems to be more closely related to seasonal influences rather than influent quality. During periods of oxygen stress in August and mid winter, BOD_5 removal was diminished.

The performance of the marsh with respect to the suspended solids parameter was variable. At Listowel marshes were shown to be excellent settling and filtering mechanisms. Yet in the Cobalt data high levels of SS were reported in the effluent on some occasions. One must be cautious in interpreting the significance of these observations because marshes are charaterized by autochothonous production of detritis. Many of the high SS discharges were coincident with adjustments in the water level. Such disturbance would have been sufficient to resuspend detrital material.

As found in Listowel phosphorus concentrations leaving the marsh during the growing season are very low, although, during the cold oxygen stressed periods of winter much of this treatment is lost.

There are no comparable objectives for secondary sewage treatment with which to evaluate the bacteriological performance of the marsh system. The marsh effluent varies widely in quality through different seasons and conditions. It can be concluded however that the bacteriological quality of the effluent in the better operating periods far exceeds the quality to be expected for a traditional secondary treatment plant.

CONCLUSION

On the basis of this discussion it seems reasonable to conclude the marsh sewage treatment technology can be transferred successfully to northern climates. The results of this study compare favourably to secondary sewage treatment standards despite the marsh being inadvertently loaded at 3 to 5 times the intended rate.

This accidental overloading provides a bonus of information to the conclusion of the study. The original design loadings were set at Listowel arbitrarily based on conservative estimates of what the marsh could handle. There has been no attempt at Listowel to define the upper loading limit the marsh system could treat.

In this study we see that higher loadings of this particular effluent can be satisfactorily treated. We cannot strictly define what these values are due to the problems with flow regulation and exfiltration. However, it seems reasonable to conclude that areal loadings of twice to three times the Listowel rate are quite acceptable at Cobalt.

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A RATIONAL METHOD FOR SLUDGE DEWATERING VIA FREEZING

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INTRODUCTION

The traditional method for sludge dewatering for many moderate to small sized wastewater treatment systems has been a simple sand drying bed. However, the increasing use of alum and other chemicals in wastewater treatment significantly affects the drainability of the resulting biochemical sludges. It may be necessary to abandon the drying beds and use more complex mechanical techniques unless a simple alternative for sludge conditioning can be developed.

Freeze thawing of sludge is one possibility. It has already been well established that freezing a sludge will convert a material with an undrainable jelly-like consistancy to a granular type material that drains immediately upon thawing. It was the purpose of the work described in this paper to determine if a rational and practical method could be developed for sludge freezing. If successful, the method should then find application as a low cost conditioning/dewatering technique for both water treatment and wastewater sludges.

BACKGROUND

The engineering literature has contained articles and descriptions of sludge freezing for at least the past 55 years (1,2,3,4,5,6). Most of these papers provide very graphic, and similar descriptions of the phenomenon and the potential benefits to be realized. Most of these papers only speculate on the physics and mechanics of the process, and a few attempt to develop a rational and practical method for implementation.

Many early, and some recent investigators suggest that compressive forces caused by ice expansion are the major factor responsible for transformation of the solid sludge particles. Other research suggests that water migration, in liquid films is the principal factor, with compression making some contribution. It has been suggested that the dewatering effect is related to: temperature, time, initial moisture content, freezing rate, electrolytical concentration, zeta potential, and chemical composition of the sludge (9). Fortunately, it is not necessary to define the interrelationship of all of these factors to attempt the development of a rational procedure for freeze dewatering.

All of the prior investigations agree that the energy costs for artificial freeze-thawing would be prohibitive so that the concept must

depend on natural freezing to be cost effective. This led to the implication that the method could only be feasible in locations with long, extremely cold winters and in fact the actual use of freezing is still essentially limited to those areas.

A few authors have suggested techniques and procedures. Fulton (7) suggested that water plant alum sludges be applied in thin layers to facilitate freezing. Rush and Stickney (6) have developed a preliminary design procedure based in essence on the freezing index for a particular location. Schleppenbach (8) has described sludge freezing in layers at the Duluth, MN, water treatment plant and indicated a relationship between the air freezing index and the frozen depth of sludge. However, in several other locations the approach seems to have been to dig a hole, fill it with sludge and hope it all freezes. The successful experiences that have been documented, demonstrate that a rational design and management procedure should be possible.

THEORY

The environmental factors that would control the freezing of sludge in a particular location are the same factors that would govern the depth of frost penetration in soil and/or the formation of ice on ponds and lakes. The depth of freezing or thawing in these latter cases can be described with Neumann or Stefan based solutions which have the basic form:

$$X = m\,[I]^{1/2} \qquad (1)$$

where: X = depth of freezing or thawing, cm
m = proportionality coefficient $(°C\text{-}d)^{-1/2}$
I = freezing (I_f) or thawing (I_t) index °C-d

The solution becomes increasingly complex for multiple layer, multiphase systems such a snow covered ice layer on a pond or several different soil layers in the profile, each with a different moisture content. However, the basic form had been used for many years to estimate the depth of ice that will form on water bodies. The coefficient m (in metric units) for these cases ranges from 1.70 for small streams or a heavy snow cover on top of the ice to 3.07 for theoretically ideal condition for freezing.

It can be shown with equation 1 that the rate of ice growth decreases with time, under steady-state conditions. This is because the ice layer itself acts as an insulating barrier between the cold ambient air and the remaining unfrozen water. It can also be shown with equation 1 that it is possible to freeze a greater total depth of ice in a given time if the water is applied in thin layers at the surface as compared to continuous ice formation on a deep body of water. This is well recognized in practice and is the basis for construction of ice bridges, hockey rinks and skating ponds. For example, assuming an average air temperature of -10°C and an "m" of 2.00, equation 1 would predict the growth of 60 cm of ice in 90 days. However, if cold ($\sim$ 0°C) water were sequentially applied in 4 cm

deep layers at the surface, about 400 cm of ice could in theory be frozen in the same period under the same conditions.

It is critical to the success of the operation with sludge that the entire layer freeze completely if the dewatering benefits are to be realized. In very cold climates with prolonged winters, the layer thickness may not be critical. However, in more temperate areas, and in particular those that experience alternating freeze-thaw periods, the depth of the sludge layer is very important. In these situations, a large single layer may never freeze completely with only the upper portion going through alternating freezing and thawing cycles.

In general, engineering designs are based on "worst case conditions" to insure successful performance of the process at all times. The typical engineering concerns with the freezing index for a particular location are usually related to the maximum depth of frost penetration, so the "worst-case" condition would be the coldest winter during the period of record. However, in this case, the focus is reversed. If sludge freezing is to be a reliable expectation every year the design must be based on the warmest winter during the period of concern, and on a layer thickness that will freeze within a reasonable time if freeze-thaw cycles occur during the winter.

Once a committment is made to sludge freezing in thin layers, other concerns arise regarding the responses and reactions during the thawing period, particularly with respect to biochemical sludges from wastewater treatment processes. Chemical sludges from water treatment are relatively inert, odor free, and have been shown to retain their desirable granular characteristics upon thawing.

The thawing responses of a frozen, layered mass of wastewater sludge had not been defined prior to this study. Since thawing will commence from the surface downward and the still frozen material beneath is relatively impermeable, will melt water pond on the surface? Will the ponded water lead to odors? Will the biochemical sludges contained in the ponded water revert to their flocculant condition and prevent effective drainage? What effect would subsequent rainfall have on an exposed sludge bed? In addition to developing a practical freezing procedure this study was designed to answer these critical questions regarding thawing.

MATERIALS AND METHODS

The experiments included small scale laboratory tests under controlled conditions and full scale outdoor field trials.

The first goal of the laboratory work was to confirm the applicability of equation 1 for freezing of typical sludge mixtures (3 to 7 percent solids) by comparing the freezing rate of the sludge mixture to the rate for water under the same conditions. A special insulated container was fabricated for this purpose. The central chamber, with a capacity of about 2L, was exposed at the top but surrounded on the sides and bottom by

insulation. The container and the liquid sample (sludge or water) were cooled to about 0.5°C and then placed in a cold room maintained at about -7°C. Data collection commenced with the first ice formation and typically continued until six to eight centimeters of ice had formed (36-48 hours). Preliminary runs confirmed that ice formed from the top and not from the sides or bottom during that period.

Another laboratory experiment froze sludge in layers in a tall plastic column. This was then insulated on the sides and bottom and thawing induced at the exposed upper surface. The purpose was to observe the effects of the ponded water and the responses of the thawing solid particles. The experiment did not succeed since a thin annular space melted around the frozen core and water drained rapidly away leaving the granular solids. The sludge used in all of these laboratory experiments was digested secondary sludge containing a significant fraction of alum.

The field trials were conducted on the surface of a large scale (9x9 m) outdoor lysimeter containing a sandy loam soil. The lysimeter has concrete walls and with the contained sandy soil is a valid representation of a typical sludge drying bed. Three sequential layers (each about 8 cm) of digested primary sludge (6 to 8% solids) were applied to the bed. Thermocouples were used to monitor temperatures and depth of freezing was measured directly. Each layer was allowed to freeze prior to the next applicaton. Observations continued through the thaw period and included several rainfall events.

RESULTS

The laboratory scale freezing experiments confirmed that sludges at concentrations typically found in the field (3 to 7% solids) freeze at essentially the same rate as tap water and that the rate can be described with equation 1 as modified below:

$$X = 1.87 \sqrt{(\Delta T)(t)}$$

where: $(\Delta T)(t)$ = freezing index, °C-d

The coefficient 1.87 derived from the experiments is within the range previously reported but may be unique for the container or the particular environmental conditions and should not be used for design. The frozen sludge samples from these experiments drained rapidly during the thawing process. Solid concentrations were approaching 25 percent at the point when thawing was complete. Comparative experiments with the same sludge indicated that the unfrozen material lost moisture via dessication and took many weeks to reach the solids concentration achieved in one day by freeze-thaw conditioning. Similar results have been observed by many prior investigators.

The first sludge at 8% solids was applied to the field test on 29 February 1984 in a layer about 10 cm deep. The sludge was quite warm (30°C) having been hauled directly in a tank truck from the treatment plant

digestor. The temperature at mid depth had cooled to about 9°C in four hours, and ice formation began within 24 hours. Probing around the perimeter indicated it was completely frozen at many points within 48 hours. The soil surface in the bed was irregular and there may have been deeper pockets near the center still not completely frozen at that point. The air temperature during this period ranged from -4°C to -17°C with negligible wind.

The second sludge application, at 7% solids, was applied in an 8 cm layer, about midday on 2 March 1984 on the frozen surface of the previous layer. The sludge was again about 30°C and the air temperature at the time of application was -4°C. The sludge temperature at mid depth in the new layer cooled to about 0°C in about 8 hours with ice formation again commencing on the surface within 24 hours. Probing near the bed perimeter again indicated complete freezing within 48 hours. Approximately five centimeters of snow fell on the bed in the next 24 hours.

The final 8 cm application, at 6% solids, was not made until 12 March 1984, only because sludge was not available until that date. The air temperature at the time of application was -6°C with a wind speed at about 16 km/hr. The initial sludge temperature was again 30°C and the application of warm sludge melted all of the snow that had accumulated on the bed. A snow storm commenced about 20 hours after the sludge was applied. The newly frozen sludge was about 3 cm thick at this time. It continued snowing for 24 hours with a total accumulation of 60 cm. Probing, 60 hours after the sludge application indicated about 6 cm of the new layer had frozen, with 1 cm of snow on top of the ice layer in a wet slushy condition. Ambient air temperatures during this 60 hour period ranged from -6°C to -15°C.

The thaw period began almost immediately after the snow storm so by the 20th of March (8 days after final application), daytime air temperatures were above 0°C. By the 23rd of March, the snow had all disappeared and there was about 6 cm of mixed meltwater and thawed sludge on top of the remaining frozen material. On the 29th of March, only 7 cm of partially frozen material remained with no ponded water on the surface and no odors. Apparently the melt water drained away through cracks in the still frozen material. The solids concentration of the thawed sludge cake on top was 26 percent. On the 2nd of April, all of the sludge was thawed and the solids concentration was 35 percent.

DISCUSSION

The total depth of liquid sludge applied was 26 cm. Allowing for about 20% expansion during freezing, the depth of frozen material would have been about 31 cm (not including the snow). This material thawed completely in 14 days with ambient air temperatures in the period ranging from 1°C to 12°C. On the 13th of April (11 days after thawing was complete), the solids concentration was about 54%. During the period 14-16 April, a steady rainfall occurred with a total of 5 cm of precipitation. There was no ponding during the rainfall and the solids concentration was

40 percent about 12 hours after the rain had stopped. Odors were never noted at any time in the thawing or drying period.

The data analysis for the various freezing and thawing observations indicated a range of values for the coefficient in equation 1, from 2.01 to 2.14. A median value of 2.04 is suggested for design, and was confirmed by independent freezing observations at a U.S. Army Depot in Pennsylvania during February-March 1984. It also can describe the freezing results reported at the Duluth, MN (8) water treatment plant.

The sludge layers used in the field trials were not specifically designed but were the result of the tank truck capacity and the bed area. A separate analysis was conducted to determine a generally applicable layer thickness for design purposes. It is unreasonable, and not cost effective to expect treatment plant operators to apply a few millimeters of sludge on a frequent schedule. On the other hand, a very thick deposit may never freeze to the bottom. Calculations with equation 1 and the coefficient determined above tend to converge on 8 cm as a practical layer for all locations. At -5°C an 8 cm layer should freeze in about 3 days, at -1°C it would take about two weeks which is still practical. A 10 cm layer, for example, would require 22 days to freeze at -1°C which is too long when the potential for freeze-thaw cycles are a factor.

A greater depth should be feasible in colder climates. Duluth, MN successfully freezes sludge in 23 cm layers. However, 8 cm should be feasible in even moderately cold climates. It is suggested that 8 cm be used for feasibility analysis and preliminary design. A larger increment may then be justified by a detailed evaluation during final design.

The sludge used in the field trial came directly from the digestor, it was very warm and required almost 24 hours to cool enough for ice formation to commence. In many other situations the sludge may come from storage and be colder but it is suggested for all situations that a one day allowance be made for sludge application and cooling in the design calculations (equation 1 is based on the assumption that the liquid at the freezing front is at 0°C).

DESIGN PROCEDURE

As indicated previously the design must be based on the warmest winter of record or period of concern to insure reliable performance at all times. The most accurate, and most cumbersome approach is to examine the weather records for a particular location and determine how many 8 cm layers of sludge could be frozen in each winter. The winter with the lowest total depth is then the design year. This approach might assume, for example, the first 8 cm application on 1 November. Equation 2, below can then be used to calculate the number of days required to freeze the layer under the average daily temperature conditions indicated in the records

$$t = \frac{\left(\frac{X}{m}\right)^2}{\Delta T} \qquad (2)$$

with 8 cm layer and m = 2.04:

$$t = \frac{15.38}{\Delta T}$$

where: t = time to freeze an 8 cm layer, d
$\Delta T = 0° - T_A$
T_A = mean daily air temperature, °C

Account is taken in the calculation of thawing periods and a new application is not allowed until the previous layer has frozen completely. One day is then allowed for application and cooling of the next layer and then equation 2 repeated to again determine freezing time. The first use of this approach was at a U.S. Army depot in eastern Pennsylvania. The calculations showed that about nine 8 cm layers of sludge could be frozen in the design year. Previously it had been suggested that freezing was not a viable approach for this location because of the relatively moderate winters.

Another, more rapid approach for preliminary design has been developed by relating the potential total depth of sludge that could be frozen to the maximum depth of frost penetration for the same location. It seemed reasonable that the two should be related since they are both dependent on the same enviromental factors. Weather records were obtained for selected locations for the period 1972-1983. Calculations were then made as described above to determine the number of 8 cm increments that could be frozen in the warmest winter during the study period. These values were then compared to the published data (10) for maximum frost penetration for the same location. The results are summarized below.

Location	Total sludge depth (Σ of 8 cm increments) cm	Maximum depth frost penetration cm
Tobyhanna, PA	72	113
Pittsburgh, PA	99	97
Denver, CO	68	107
Bismark, ND	229	201
Pocatello, ID	91	102
Duluth, MN	290	206

The equation describing these data takes the form:

$$\Sigma X = 1.76\ (F_p) - 101 \qquad (3)$$

where: ΣX = total depth of sludge that could be frozen in 8 cm increments, cm
F_p = maximum depth of frost penetration

the r^2 for this correlation is 0.92.

As indicated by equation 3 sludge freezing will not be feasible unless the depth of maximum frost penetration is greater than 57 cm for a particular location. In general, that will begin to occur above the 38th parallel and include most of the northern half of the United States, with the exception of the west coast. However, it is unlikely that sludge freezing will be cost effective, because of the large area required, if only one or two 8 cm layers can be frozen in the design winter. A frost penetration of about 100 cm would allow sludge freezing for a total of 75 cm and depending on land and construction costs, the process may be cost effective at that stage. Frost penetration data can be found in reference 10; representative values and the calculated sludge depths are given below.

Location	Maximum frost penetration cm	Depth of sludge freezing, calculated with equation 3 cm
Bangor, ME	183	221
Concord, NH	152	166
Hartford, CT	124	117
Chicago, IL	122	113
Omaha, NB	114	100
Minneapolis, MN	190	233
Rapid City, SD	162	184
Butte, MT	127	122
Montreal, Que.	203	256

Example

A community near Chicago, IL is considering sludge freezing as a dewatering technique for their annual 3000 m^3 sludge production.

$$\text{Area for freezing-drying beds} = \frac{3000\ m^3}{\Sigma X} = \frac{3000}{1.13} = 2654\ m^2$$

Could use 16 beds, each 7 m x 24 m
Allow 20% for expansion, and 16 cm freeboard, so:
Depth: $(1.20)(1.13) + .16 = 1.5$ m

It is suggested that the drained biochemical sludges be removed each year. Inert chemical sludges from water treatment could remain in place for several years. In these cases, a bed 2 to 3 m deep could be constructed and the residual sludge solids allowed to accumulate for several years prior to removal and disposal. The bed details can be designed with standard drying bed criteria. The final design should evaluate actual weather records to determine the "worst-case" year.

CONCLUSIONS

Sludge freezing can be a reliable dewatering method for most of the northern United States and Canada. The rational procedure described in

this paper can be used for design and for management by plant operators. The procedure can be used for design of new systems as well as conversion of existing drying beds. Exposed beds will be more effective than covered units, however, the freezing process will be accelerated if accumulated snow is removed. Cost effectiveness of the process will depend on area requirements and land costs and on the capability to store sludge during the warm part of the year. Odors should not be a problem with digested biochemical sludges. Odors may be a concern during the thawing stage with unstabilized biological sludges.

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ACKNOWLEDGMENTS

The work described in this paper was supported by U.S. Department of the Army Project 4A762730AT42. The authors wish to acknowledge Ms. Diane Seely for her computational assistance and correlation analysis, Mr. Carl Deiner, Don Chamberlain, and Carl Elder for their assistance in the field tests; and the town of Hanover, N.H., for the sludge.

ANAEROBIC TREATMENT OF HIGH STRENGTH ACIDIC ORGANIC WASTEWATERS UTILIZING THE UPFLOW SLUDGE BLANKET TREATMENT PROCESS

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ABSTRACT

A laboratory bench-scale study was undertaken in order to investigate the anaerobic biological treatment of high-strength acidic organic wastewaters utilizing the Upflow Sludge Blanket (USB) treatment process. A synthetic wastewater was utilized having a pH of 4.2 and consisting primarily of acetic and propionic acids.

While operating at substrate loading rates of 1.1 and 2.4 kg COD/kg VSS/d (e.g., 10 and 30 kg COD/m^3/d, respectively), the USB treatment process removed in excess of 90 per cent of the total COD present in the raw wastewater, for waste strengths of 2 to 32g COD/L and hydraulic retention times varying between 3.2 and 76.8 hours. The process demonstrated the ability to polish the effluent waste stream to effluent COD concentrations of less than 300 mg/L. The process was able to tolerate hydraulic and waste strength shockloads. No observable deterioration in pH or effluent quality was identified following sudden increases in wastewater concentration (e.g., 4 to 32g COD/L) or decreases in hydraulic retention time (e.g., 25.6 to 3.2 hours). The use of effluent recirculation did not influence treatment efficiency but was necessary in order to partially neutralize the influent low pH wastewater.

The results of this research also demonstrate the feasibility of utilizing the USB reactor as the methane forming reactor in a two-stage anaerobic wastewater treatment system. Recommendations are provided concerning the design of such a system.

INTRODUCTION

The anaerobic biological treatment process, commonly referred to as anaerobic digestion, was traditionally utilized almost exclusively for the stabilization of sewage sludge. The process received little application in the treatment of organic industrial wastes due to several limitations associated with the process, including the low achievable rates of performance, the inability to withstand hydraulic or organic shockloads and the poor process control (1, 2). These problems, all inherent to conventional digesters, were attributed to difficulties in retaining biomass (solids) within the digester. However, rising energy costs have spurred the development of anaerobic waste treatment technology and more specifically, the design of reactors or process

configurations capable of achieving high solids retention times (3). One such reactor, the Upflow Sludge Blanket (USB) reactor, relies on the development of sludge exhibiting superior settling characteristics, flocculating with time to form tightly-bound aggregates of micro-organisms, which are commonly referred to as sludge granules (4). It operates entirely as a suspended growth system and uses no packing material to support and concentrate biological growth.

Although the USB reactor has been shown to be effective in treating a wide variety of wastewaters, its performance remains dependent upon the presence of "granular sludge" within the reactor (1,2,3). In turn, the development of such sludge appears to be waste dependent. Previous studies have identified that the treatment of wastewaters containing a high fraction of finely dispersed suspended solids, or composed of refractory organic compounds, did not result in the development of sludge granules, whereas the treatment of dilute volatile fatty acid mixtures produced distinct highly active sludge granules (5,6,7). Volatile acid mixtures are commonly found in a two-stage anaerobic treatment process (e.g. hydrolysis/acidification phase followed by methane-forming phase), specifically in the effluent stream from the acid-forming reactor and introduced into the methane- forming reactor.

In light of these findings, an investigation was undertaken in order to identify the suitability of the USB reactor for the treatment of acidic wastewaters (e.g. containing a high concentration of volatile acids) and more particularly, as the methane phase reactor of a two-stage treatment process. During the first stage of this investigation, process start-up was investigated and results have been previously reported elsewhere (8). It was found that the use of a more active, densely flocculated anaerobic sludge, as opposed to sewage sludge originating from a secondary digester, allowed for the development of tightly-bound clusters of biological solids (i.e. sludge granules) and correspondingly a more rapid start-up. Within 25 days of start-up, the sludge present in the bed of these reactors was entirely of a granular or pelletized form, with granules measuring up to 2 mm in diameter. Microscopic examinations of the sludge granules identified that the granules were composed mainly of filamentous micro-organisms.

This paper reports the second stage of this investigation. Specific objectives were to evaluate the influence of certain operating parameters, such as the wastewater concentration, the hydraulic retention time and the effluent recycle ratio at fixed volumetric substrate loading rates, on the process' treatment efficiency and stability.

MATERIALS AND METHODS

Synthetic Waste and Inoculum

In order to simulate a high-strength acidic organic wastewater, use was made of a synthetic wastewater consisting primarily of acetic and propionic acids and enriched with nitrogen, phosphorus, sulphur, trace metals and yeast extract to sustain microbial growth. In addition, calcium ion in the form of $CaCl_2$ was maintained in the feed waste, based on previous identifications which identified the positive influence of divalent cations on the development of sludge granules (9). The components of the feed solution and the characteristics of this waste are outlined in Tables 1 and 2 respectively.

TABLE 1: FEED SOLUTION COMPONENTS AND CONCENTRATIONS

Acetic acid	5000 ± 100 mg/L
Propionic acid	1000 ± 75 mg/L
Yeast extract	750 mg/L
KH_2PO_4	200 mg/L
NH_4NO_3	70 mg/L
$(NH_4)_2SO_4$	20 mg/L
$CaCl_2$	80 mg/L
*Trace Metal Solution	2.0 mg/L
NH_4HCO_3	650 mg/L
$NaHCO_3$	2250 mg/L

*Trace metal Solution (10)

$FeCl_2 \cdot 4H_2O$	2000 mg/L
H_3BO_3	50 mg/L
$ZnCl_2$	50 mg/L
$CuCl_2 \cdot 2H_2O$	40 mg/L
$MnCl_2 \cdot 4H_2O$	500 mg/L
$(NH_4)_6Mo_7O_{24} \cdot 4H_2O$	50 mg/L
$AlCl_3$	30 mg/L
**$CoCl_2 \cdot 6H_2O$	150 mg/L
**$NiCl_2 \cdot 6H_2O$	100 mg/L
EDTA	500 mg/L
Concentrated HCl	1 mL/L

**Modified according to Murray and van den Berg (11).

TABLE 2: CHARACTERISTICS OF THE FEED SOLUTION

Total COD	8000 + 300 mg/L
*VFA - COD	6810 + 220 mg/L
pH	4.6
Alkalinity	1600 mg/L, as $CaCO_3$
Total Nitrogen	190 mg/L
NH_3 - Nitrogen	110 mg/L
Phosphorus (total)	45 mg/L
Potassium	57 mg/L

COD : N : P
= 7700 : 190 : 45
= 171 : 4.2 : 1.0

COD : N
= 7700 : 190
= 40 : 1

*Volatile Fatty Acid (VFA) COD - based on 1.065g COD/g acetic acid and 1.513g COD/g propionic acid.

Such a waste stream may be representative of easily-hydrolyzed soluble organic wastewaters. More specifically, a high concentration of these volatile organic acids are usually found in such wastewaters at their point of discharge, or shortly after discharge. These acids are rapidly produced from the dissolved biodegradable compounds. It is also representative of the acidic waste stream emanating from an acid-phase reactor in a two-stage anaerobic treatment process. Specifically, when the acid-phase digestion process is operating at steady-state sub-optimal rates, acetic acid is the most prevalent and propionic acid, the most refractory of the acidic reaction products (12,13).

All feed solutions were prepared with tap water. They were prepared in advance in 150L volumes and stored at 4°C in a refrigerated room until required. The feed solution described in Tables 1 and 2 was utilized as a stock solution and diluted when wastes of identical composition but of lower concentrations were required. For wastes of a higher concentration, the components of the feed mixture were increased accordingly.

Three identical USB reactors were innoculated with sludge obtained from an inoperative anaerobic reactor previously utilized to treat a food processing waste. This reactor had remained inoperative for a period of six months. This sludge was not granular or pelletized but rather was of a very dense flocculant nature. It was characterized by a specific microbial activity of approximately 0.47 kg COD/kg VSS/d, and a total solids content of 23.6g VSS/L. Each reactor was inoculated with 1.5L of sludge and the remaining volume was filled with tap water.

Apparatus

The USB reactors utilized have been described previously (14). A schematic of the system utilized is illustrated on Figure 1. The system consisted of a 4.2L reactor (column height 80 cm, diameter 9.4 cm) and an external solids settling chamber (modified inverted Erlenmeyer flask) of 2.2L in volume. The reactors were placed in a walk-in temperature controlled room maintained at 35 $\pm$ 1°C. A refrigerated feed container was utilized to keep the feed solution cool (4°C). The recycled effluent stream merged with the influent raw feed prior to the reactor inlet, ensuring that the influent waste was introduced into the reactor at temperatures near 35°C.

A dual channel Harvard Peristaltic Pump Model 1203 served as both a feed and recycle pump. All fluid and gas lines were of Tygon tubing.

Analytical Methods

Gas composition was measured by gas chromatography in accordance with the method of van Huysteen (15) using a Hewlett-Packard 5710A gas chromatograph. Measurement of gas volumes was accomplished utilizing a wet test gas meter (Precision Scientific Model 6311). Volumes of methane were corrected to standard temperature and pressure.

Volatile acid concentrations were measured by gas chromatography in accordance with the method of Ackman (16), using a Hewlett-Packard gas chromatograph Model 5840A equipped with automatic sampler Model 7672A and integrator Model 3380A.

Measurement of solids was performed in accordance with the methodology outlined in Standard Methods (APHA 1975). pH was measured with a glass electrode, using a Radiometer/Copenhagen pH Meter Model 26.

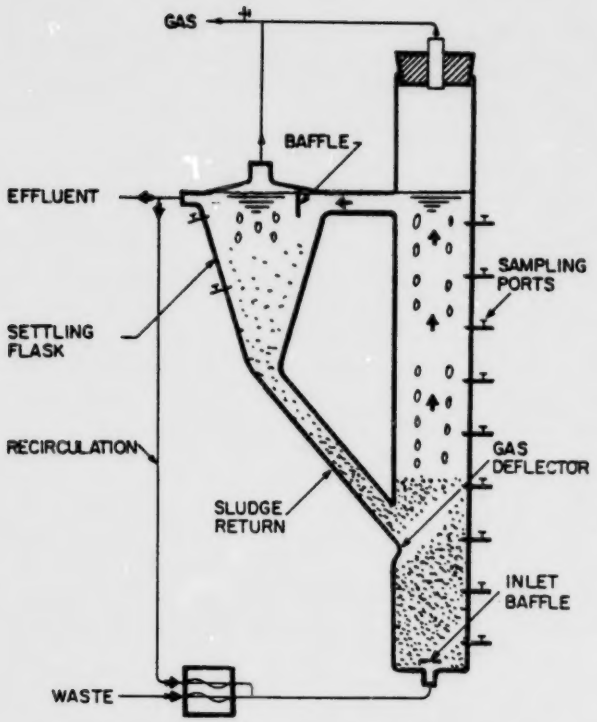


FIGURE 1: UPFLOW SLUDGE BLANKET (USB) REACTOR SYSTEM

Chemical oxygen demand (COD) determinations were made in accordance with the methodology outlined by Knetchel (17). Both the total and soluble COD of the effluent were evaluated. In order to measure the soluble COD, the sample was centrifuged at 6,000 rpm for 20 minutes, after which the COD analysis was performed on the supernatant.

Microbial activity (e.g. specific activity) was measured following the method of Murray and van den Berg (11). Samples were withdrawn via the sampling points located along the height of the reactor and were transferred anaerobically into serum bottles. Both the acetic acid degradation rate and the propionic acid degradation rate were assessed in order to define the microbial activity of the sludge.

EXPERIMENTAL PROGRAM

The neutralization requirements for the anaerobic treatment of this low pH acidic wastewater were not investigated. Alternatively, use was made of the recirculation of effluent to buffer this wastewater by adding biologically produced alkalinity to the influent and to thus reduce the requirement for neutralization chemicals.

Three reactors having previously been inoculated with 1.5L of anaerobic sludge were operated in parallel. A waste concentration of 8g COD/L and an effluent to feed recycle ratio of 4 were utilized in systematically

increasing the loading rates of all three reactors to 10 kg COD/m^3/d. Sludge found in all three reactors at the time of this investigation was of a granular form. Results and discussion of the start-up of the process and development of sludge granules are described elsewhere (8).

While operating at a volumetric organic loading rate of 10 kg COD/m^3/d and a recycle ratio of 4, two USB reactors were subjected to raw wastewater concentrations of 2, 4, 8, 16 and 32 g COD/L. These waste strengths were investigated in a random order in the reactors so as to avoid the possibility of gradual adaptation of the sludge to higher waste strengths. A minimum of 4 days or 6 hydraulic retention times (HRT) was allowed for each concentration investigated. The concentration of volatile fatty acids (VFA) in the effluent at the time of discharge was sampled for and immediately analyzed two times per day. The total and soluble COD in composite samples of effluent (i.e. samples representing the total effluent discharged over a 24-hour period) were sampled for at least three times over the course of the test run, for each wastewater concentration investigated. Samples were refrigerated at 4°C until analyzed. Gas production and composition were analyzed daily.

Following the completion of these investigations, the volumetric organic loading rate was gradually increased in a stepwise manner to a rate of 30 kg COD/m^3/d using a wastewater having a concentration of 8g COD/L and a recycle ratio of 4. After ensuring that steady state conditions prevailed, the impact of raw waste concentrations (i.e. 4, 8, 16, 32g COD/L) on the treatment conditions was examined. Procedure for sampling and analysis were as described above. The tests were subsequently repeated at recycle ratios of 2 and 1 in order to further assess the impact of greater volatile acid concentrations or waste strengths (e.g., less diluted) introduced directly into the reactor on the process' treatment efficiency and stability.

A third USB reactor was utilized in conducting further testing on the influence of effluent recirculation on treatment efficiency and process stability. While operating at a substrate loading rate of 30 kg COD/m^3/d (3.2 kg COD/kg VSS/d), the treatment process was consecutively subjected to recycled effluent to feed ratios of 12, 8, 4, 2, 1 and no recycle while treating a wastewater containing 32g COD/L. A minimum of 6 HRT were allowed for each recycle ratio investigated. Sampling and analyses were conducted as described earlier.

RESULTS AND DISCUSSION

Influence of Wastewater Strength and HRT

The influence of the various acidic wastewater strengths and HRT's investigated, on the treatment efficiency, is illustrated on Table 3 and Figures 2, 3 and 4. These results demonstrate the ability of the USB treatment process to effectively treat acidic organic wastewater concentrations ranging from 2g COD/L to 32g COD/L and HRT varying between 3.2 and 76.8 hours, while operating at specific substrate loading rates of approximately 1.1 and 2.4 kg COD/kg VSS/d (i.e., 0.9-2.0 kg VFA-COD/kg VSS/d). Excellent total, soluble and VFA-COD removal efficiencies (e.g., percentage of raw wastewater COD removed), each exceeding 92 per cent, were observed. The USB treatment process exhibited the ability to not only purify the influent wastewater but also to "polish" the effluent waste stream to effluent VFA-COD

TABLE 3: SUMMARY OF RESULTS - WASTE STRENGTH TREATMENT STUDY

Volumetric Substrate Loading Rate ($kgCOD/m^3/d$)				Recycled Effluent to Feed Ratio	Raw Waste Concentration (mg/ℓ)		Diluted Influent[2] Waste Concentration (mg/ℓ)		Hydraulic Retention Time (hrs)		Hydraulic Surface Loading Rate	Influent pH	Operating pH	Effluent Waste Concentration (mg/ℓ)			Waste Removal Efficiencies (% of raw waste removed)			Substrate Removal Rate ($kgCOD/m^3/d$)				Specific Substrate Removal Rate ($kgCOD/kgVSS/d$)	
Total Volume[1]		Reactor Unit																		Total Volume[1]		Reactor Unit			
Total COD	VFA COD[3]	Total COD	VFA COD[3]		Total COD	VFA COD[3]	Total COD	Total[3] VFA-COD	Total Volume[1]	Reactor Unit				Total COD	Soluble COD	VFA COD[3]	Total COD	Soluble COD	VFA COD[3]	Soluble COD	VFA COD	Soluble COD	VFA COD	Soluble COD	VFA COD[3]
10	8.5	15.2	12.9	4	2000	1700	504	442	4.8	3.1	23.05	6.4	6.8	135	100	135	93.3	95.0	92.1	9.5	7.8	14.5	11.9	1.04	0.86
					4000	3400	924	737	9.6	6.3	11.53	6.4	6.9	160	95	75	96.1	97.7	97.9	9.8	8.3	14.9	12.6	1.07	0.91
					8000	6800	1756	1406	19.2	12.6	5.77	6.3	7.0	180	95	70	97.7	98.8	99.0	9.8	8.4	15.0	12.8	1.07	0.92
					16000	13600	3384	2760	38.4	25.2	2.88	6.4	7.3	230	150	55	98.6	99.0	99.6	9.9	8.4	15.1	12.8	1.08	0.93
					32000	27200	6692	5650	76.8	50.4	1.44	6.5	7.5	365	205	130	98.9	99.3	99.5	9.9	8.4	15.1	12.8	1.08	0.93
20	25.5	45.7	38.8	4	4000	3400	934	792	3.2	2.1	34.60	6.4	6.4	190	125	140	95.2	96.9	95.9	29.1	24.4	44.3	37.2	2.34	1.96
					8000	6800	1794	1484	6.4	4.2	17.30	6.3	6.8	280	195	155	96.5	97.6	97.7	29.3	24.9	44.6	37.9	2.34	1.99
					16000	13600	3364	2825	12.8	8.4	8.65	6.5	7.3	355	195	140	97.7	98.7	99.1	29.6	25.2	45.1	38.3	2.37	2.02
					32000	27200	6605	5540	25.6	16.8	4.32	6.5	7.5	525	245	135	98.3	99.2	99.5	29.8	25.4	45.4	38.7	2.38	2.03
30	25.5	45.7	38.8	2	4000	3400	1428	1171	3.2	2.1	20.76	5.8	6.5	145	70	60	96.4	98.2	98.3	29.5	25.1	44.9	38.2	2.36	2.00
					8000	6800	2808	2333	6.4	4.2	10.38	5.7	6.8	215	125	105	97.3	98.4	98.5	29.5	25.1	44.9	38.2	2.36	2.01
					16000	13600	5617	4666	12.8	8.4	5.19	5.8	7.2	405	240	190	97.3	98.5	98.6	29.5	25.1	44.9	38.3	2.36	2.01
					32000	27200	11262	9220	25.6	16.8	2.60	5.9	7.6	955	385	260	97.0	98.7	99.0	29.6	25.2	45.1	38.4	2.37	2.02
30	25.5	45.7	38.8	1	4000	3400	2000	1700	3.2	2.1	13.84	5.1	-		Upset of Treatment Process											
					8000	6800	4000	3400	6.4	4.2	6.82	5.1	-		Upset of Treatment Process											
					16000	13600	8000	6800	12.8	8.4	3.46	5.0	-		Upset of Treatment Process											
					32000	27200	16000	13600	25.6	16.8	1.73	5.1	-		Upset of Treatment Process											
15	12.8	22.8	19.4	1	16000	13600	8000	6800	25.6	16.8	1.73	5.0	-		Upset of Treatment Process											
15	12.8	22.8	19.4	No recycle	32000	27200	32000	27200	25.6	16.8	0.43	4.5	-		Upset of Treatment Process											

1. Based on the reactor volume plus the external settler volume.

2. Influent Concentration = $\frac{C_i + RC_e}{1+R}$; where: R=effluent recirculation ratio, C_e = concentration in effluent, C_i = influent concentration (ie. undiluted waste).

3. VFA - Volatile Fatty Acids, VFA-COD calculations based on 1.06gCOD/g acetic acid and 1.51gCOD/g propionic acid.

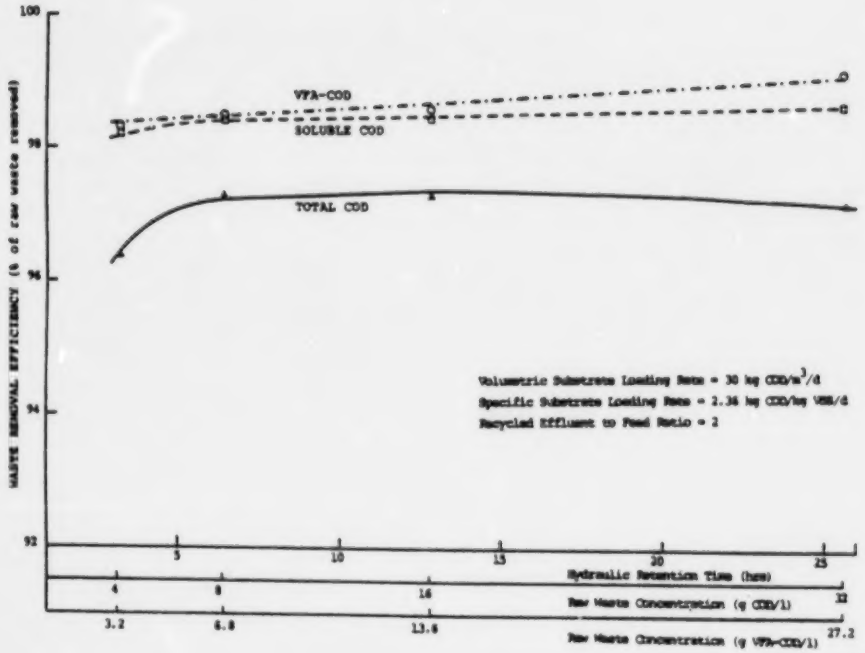


FIGURE 2: WASTE REMOVAL EFFICIENCY VERSUS HYDRAULIC RETENTION TIME AND RAW WASTE CONCENTRATION

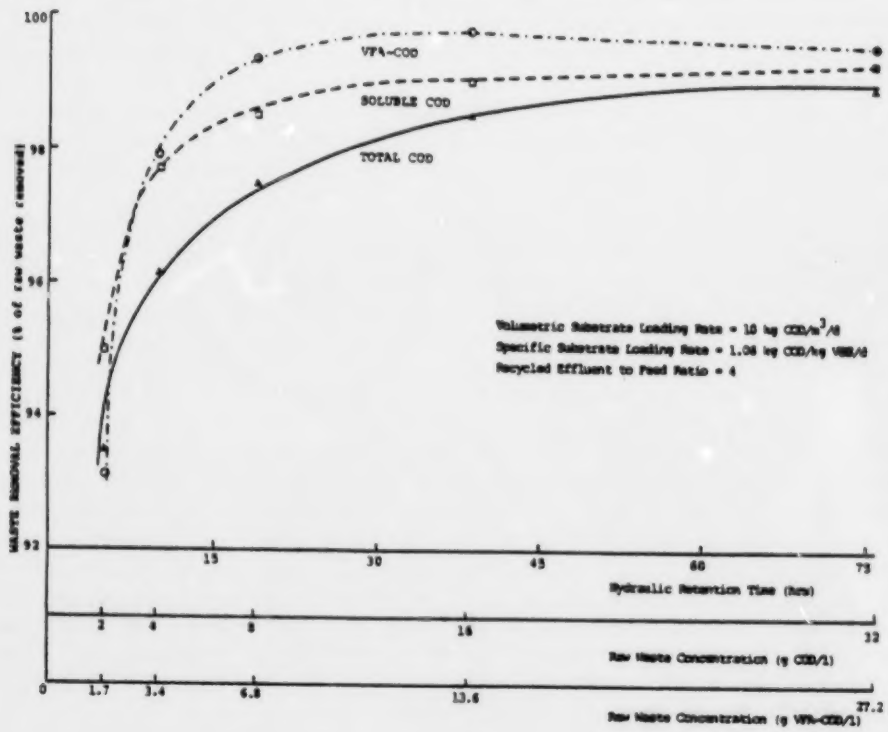


FIGURE 3: WASTE REMOVAL EFFICIENCY VERSUS HYDRAULIC RETENTION TIME AND RAW WASTE CONCENTRATION

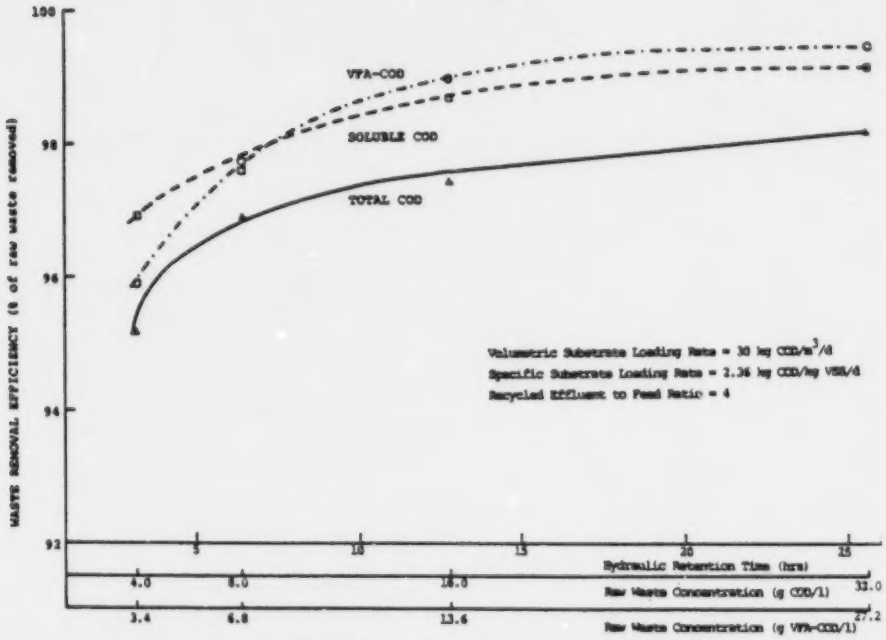


FIGURE 4: WASTE REMOVAL EFFICIENCY VERSUS HYDRAULIC RETENTION TIME AND RAW WASTE CONCENTRATION

concentrations of less than 300 mg/L. This demonstrated ability to "polish" high strength acidic wastewaters ranging up to concentrations of 27.2g VFA-COD/L, while operating at relatively high loading rates, may eliminate any requirement for secondary aerobic treatment in order to polish the effluent prior to its discharge, particularly if being discharged to a sanitary sewage system.

The removal efficiencies corresponding to the low and medium strength wastewater concentrations examined in this study compare favourably with those observed in other treatment studies operating under similar conditions. In the USB treatment of an acidic wastewater containing 2.5g VFA-COD/L, while operating at a specific substrate loading rate of 2.1 kg COD/kg VSS/d (i.e., 33.6 kg COD/m^3/d), Lettinga and co-workers (6) observed a soluble COD removal efficiency of approximately 87 per cent. When treating a soluble acidic wastewater of 13g VFA-COD/L (i.e., landfill leachate) using an anaerobic filter, Ghosh and Henry (18) obtained soluble COD reductions of 74 per cent or greater while operating at a volumetric substrate loading rate of 6.1 kg COD/m^3/d. However, in the treatment of higher strength acidic wastewaters (i.e. 26.5g VFA-COD/L), using a completely mixed anaerobic filter, Chun and de Walle found that VFA-COD removal efficiencies greater than 95 per cent were obtained only when operating at a HRT greater than 7 days (i.e. 3.7 kg VFA-COD/m^3/d). At a HRT of 3 days (i.e. 8.8 kg VFA-COD/m^3/d), VFA-COD removal efficiencies deteriorated to 72 per cent. According to Lettinga (6), this poorer treatment efficiency normally associated with low HRT's, is expected to be the result of a less effective contact between the active sludge and the substrate. In examining the influence of HRT on removal efficiency in the present study, as illustrated in Figures 2, 3 and 4, it would appear that HRT's varying between 3.2 to 76.8 hours, exerted little influence on the waste removal efficiency.

As identified on Table 3, the effluent recycle ratio was reduced from 4 to 2 and subsequently to 1, in order to further assess the impact of higher

acid concentrations or waste strengths directly introduced into the reactor on its treatment efficiency. When comparing the treatment efficiency associated with recycle ratios of 4 and 2, the HRT, substrate loading rate and raw wastewater concentration were maintained constant. Consequently, any differences in process stability or effluent quality under the conditions studied indicate the possible influence of diluted influent wastewater concentration, influent pH or the hydraulic surface loading rate. Table 3 identifies that the USB treatment process maintained a stable operation with an effluent recycle ratio of 2, despite consistent influent pH values of slightly below 6.0. In reducing the degree of effluent recirculation, the concentration of VFA introduced into the reactor was increased, the pH of the influent decreased and the hydraulic surface loading rate was decreased. As illustrated on Figure 4, both total, soluble and VFA-COD removal efficiencies were maintained well above 95 per cent in all cases investigated.

With regard to conversion end products, approximately 90-95 per cent of the theoretical methane yield of 0.35 m^3 at STP per kg COD degraded (19) was consistently obtained for all wastewater strengths examined. This finding corresponds with the observed soluble COD treatment efficiencies of 95 per cent or greater, with the balance being attributed to nominal leaks in flow lines or dissolved gas in the effluent.

Influence of Effluent Recirculation

Results from a further investigation undertaken on the impact of effluent recirculation are identified on Table 4. As shown, neither the range of hydraulic surface loading rates nor the diluted influent wastewater concentrations investigated appeared to exert any influence on the treatment efficiency. Waste removal efficiencies ranging from 96.4 to 99.1 were observed.

TABLE 4: INFLUENCE OF EFFLUENT RECIRCULATION ON TREATMENT EFFICIENCY

Volumetric Substrate Loading Rate (kg COD/m^3/d)				Recycled Effluent to Feed Ratio	Diluted Influent Waste Concentration (mg/l)[2]		Hydraulic Surface Loading Rate[3] (m^3/m^2/d)	Influent pH	Operating pH	Effluent Waste Concentration (mg/l)			Waste Removal Efficiency (% of raw waste removed)		
Total Volume[1]		Reactor Volume													
Total COD	VFA COD	Total COD	VFA COD		Total COD	VFA COD				Total COD	Soluble COD	VFA COD	Total COD	Soluble COD	VFA COD
30	25.5	45	38.5	12	3218	2092	11.25	-	7.5	820	460	390	97.4	98.5	98.6
				8	4204	3338	7.79	7.0	7.55	730	440	355	97.7	98.6	98.7
				4	6948	5632	4.32	6.5	7.6	685	320	240	97.8	99.0	99.1
				2	11267	9237	2.60	5.9	7.65	1030	395	275	96.8	98.7	99.1
				1	16515	13915	1.73	5.1	7.55	1160	530	630	96.4	98.0	97.7
				No recycle	32000	27200	0.87	4.6	-	- - - - Upset of Treatment Process - - - -					

[1] Based on reactor volume of 4.2 litres plus external settling chamber volume of 2.2 litres.

[2] Diluted influent concentration = $\frac{C_i + RC_e}{1 + R}$ where: R = recycled effluent to feed ratio
C_e = effluent waste concentration
C_i = undiluted raw wastewater concentration
= 32 g COD/l
= 27.2 g VFA-COD/l

[3] Based on recycled effluent flowrate plus raw wastewater flowrate.

Similar results were observed by Lettinga and co-workers (6) using a medium strength neutralized VFA wastewater (i.e. 5240 mg COD/L). The waste removal efficiency remained above 95 per cent for effluent recycle ratios varying from 15 to no recycle.

As illustrated on Table 3, upsets in the steady state operation of the treatment process occurred as the recirculation of effluent was reduced to 1 or completely eliminated. However, in this particular investigation an effluent recycle ratio of 1 could be tolerated although the complete elimination of effluent recirculation could not. In monitoring the upset of the treatment process, it was found in all cases that the upset was precipitated by the low influent pH coupled with an insufficient mixing or diffusion of the wastewater required to neutralize this low pH stream at its point of entry into the reactor. The pH in the reactor gradually decreased to 4.4-4.5, which corresponds to the pH of the raw wastewater, and VFA levels in the reactor increased accordingly. At a pH below 6.0, the metabolic activity of methanogenic bacteria decreases significantly in an anaerobic fermentation, resulting in the accumulation of VFA (3,20). A further description of the upset of the USB treatment process has been previously described elsewhere (8).

In contrast to observations identified on Table 3, a waste concentration of 32g COD/L with an influent pH of 5.1 could be tolerated in the present study, when operated at a recycled effluent to feed ratio of 1. This is believed to be associated with the nature of the sludge present in the USB reactors. Although the sludge in all three reactors was of a granular form, the character of the sludge in the reactor utilized in this particular recycle study was much more flocculent in nature than in the other two reactors. Such a difference in sludge character, may have been caused by a severe substrate overload experienced during the start-up of this reactor system. Previous investigations (7) have identified that diffusional resistances associated with such flocculent granules are expected to be lower than those associated with more dense or compact granules.

Influence of Shockloads - Solids and Microbial Activity Distribution Profiles

The USB treatment process also demonstrates an excellent ability to tolerate hydraulic shockloads and shocks in waste strengths. When investigating the influence of wastewater concentrations or HRT's as described previously, the process was able to adapt from one concentration to the next with no observable deterioration in pH or effluent quality, and no accumulation of VFA. Specifically, upset of the process did not occur even when a sudden large change in wastewater concentration (e.g. from 4 to 32g COD/L) or when the hydraulic retention time was reduced from 25.6 to 3.2 hours. This finding was in contrast to that of Chian and de Walle who studied the use of an anaerobic filter reactor for the treatment of high strength acidic wastewaters. They observed a deterioration in effluent quality lasting 15 days after doubling the raw wastewater concentration and the corresponding HRT.

Lettinga and co-workers noted the USB treatment process accommodated well to hydraulic and organic shockloads, provided the pH throughout the reactor remained well above 6.0 and that the specific substrate loading rate applied to the treatment process was kept below the maximum specific substrate removal rate (i.e., the maximum microbial activity). In order to assess whether these findings compare favourably with those obtained in the present study, the maximum microbial activities (e.g. acetate removal rate, propionate removal rate) in the sludge bed were evaluated and compared to the

specific substrate loading rate applied. As identified on Table 5, the maximum microbial activities in the sludge bed exceed the corresponding substrate loading rate applied during the continuous operation of the treatment process. Further, the microbial activity of the sludge was observed to increase with an increase in the substrate loading rate applied to the treatment process. Acetate removal rates in the sludge bed were found to increase from a range of 1.3-2.3 kg acetate/kg VSS/d to 2.3-2.8 kg acetate/kg VSS/d, corresponding to an increase in the substrate loading rate from approximately 0.9 to 2.0 kg acetate/kg VSS/d. Similar increases in microbial activity were observed for propionate. These observations correspond to those made by Cohen (13) when studying the kinetic responses to shockloading of USB methanogenic reactors inoculated with granular sludge. Such observations may be substantiated if it is assumed that substrate removal can be described by Michaelis - Menton kinetics. Further, the substrate degradation rates observed in the USB treatment process may also be influenced by the physical and biological characteristics of the sludge granules present in the reactor. An elaboration of these subject areas as they relate to the observations of this research can be found elsewhere (8). Their further discussion was found to be beyond the scope of this paper.

Figure 5 provides a general illustration of the distribution of the volatile suspended solids in the reactors of two of the USB treatment processes. As shown, these solids, consisting mainly of sludge granules, are concentrated in a dense bed occupying the lower 40 to 50 per cent of the reactor volume. Their shape does not reflect the dynamic behaviour of the

TABLE 5: ACETATE AND PROPIONATE REMOVAL RATES

Distance from Reactor Bottom (% of reactor volume beneath) (mm)			Solids Concentration (g VSS/)	Acetate Removal Rate Volumetric ($kg/m^3/d$)	Acetate Removal Rate Specific (kg/kg VSS/d)		Propionate Removal Rate Volumetric ($kg/m^3/d$)	Propionate Removal Rate Specific (kg/kg VSS/d)	r
100	(16)	I_a	31.2	39.9	1.28	0.998	6.2	0.20	0.986
		II_a	28.3	40.3	1.42	0.997	6.0	0.22	0.941
		I_b	36.3	102.4	2.82	0.999	13.7	0.38	0.999
		II_b	34.8	97.0	2.80	0.997	14.1	0.40	0.999
230	(38)	I_a	14.9	34.9	2.34	0.998	2.7	0.18	0.948
		II_a	13.8	30.3	2.20	0.989	2.1	0.15	0.973
		I_b	34.7	84.3	2.42	0.998	12.3	0.35	0.999
		II_b	35.9	82.5	2.30	0.997	12.7	0.35	0.999
380	(60)	I_a	0.86	1.16	-	0.987	0.25	-	0.901
		II_a	0.63	0.68	-	0.964	0.25	-	0.967
		I_b	0.83	5.2	-	0.995	0.80	-	0.993
		II_b	-	-	-	-	-	-	-
600	(96)	I_a	0.60	0.80	-	0.991	0.23	-	0.992
		II_a	0.36	0.42	-	0.960	0.25	-	0.968
		I_b	0.96	4.9	-	0.993	0.60	-	0.994
		II_b	-	-	-	-	-	-	-

I, II -Results from two separate microbial activity investigations.

a -Experiment conducted while an acetate loading rate of 0.22 kg acetate/m^3/d (0.90 kg acetate/kg VSS/d) and propionate loading rate of 1.25 kg propionate/m^3/d (0.14 kg propionate/kg VSS/d) were applied to the treatment process, with greater than 95% acetate and propionate removal efficiency.

b -Experiment conducted while an acetate loading rate of 24.75 kg acetate/m^3/d (1.98 kg acetate/kg VSS/d) and propionate loading rate of 3.75 kg propionate/m^3/d (0.30 kg propionate/kg VSS/d) were applied to the treatment process, with greater that 95% acetate and propionate removal efficiency.

r -Statistical Correlation co-efficient for the linear volumetric removal rates.

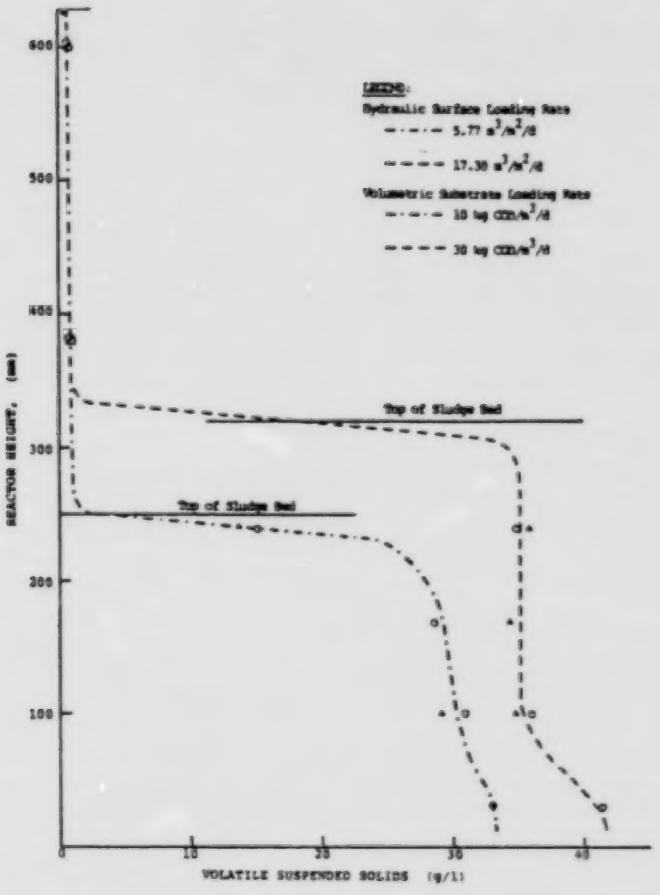


FIGURE 5: SOLIDS DISTRIBUTION PROFILES

sludge bed and represents no more than an instantaneous illustration, strongly dependent on the pattern of the hydraulic and substrate loading rates, preceding and during the sampling moment.) The volumetric acetate and propionate removal rates (i.e. the microbial activity) corresponding to these solids distribution profiles were evaluated and are tabulated in Table 5. Profiles of the volumetric distribution of microbial activity, modelled from the respective solids profiles, are illustrated in Figures 6 and 7. The sludge bed heights of 250 and 370 mm correspond to hydraulic surface loading rates of 5.77 and 17.3 $m^3/m^2/d$, respectively.

The distribution of solids and volumetric microbial activity illustrated on Figures 6 and 7 is similar to those previously observed by Lettinga (4) and co-workers and by Van den Berg and Kennedy (14). These figures substantiate that most of the COD removal occurred in the lower 30-40 per cent of the reactor where the solids are concentrated. Correspondingly, calculated volumetric substrate loading rates of 80-100 kg $COD/m^3/d$ could be accommodated in the lower 40-50 per cent of the reactor volume. Judging from these observations, it appears that the USB treatment process, while operating at the specific substrate loading rates identified on Table 5, is not yet

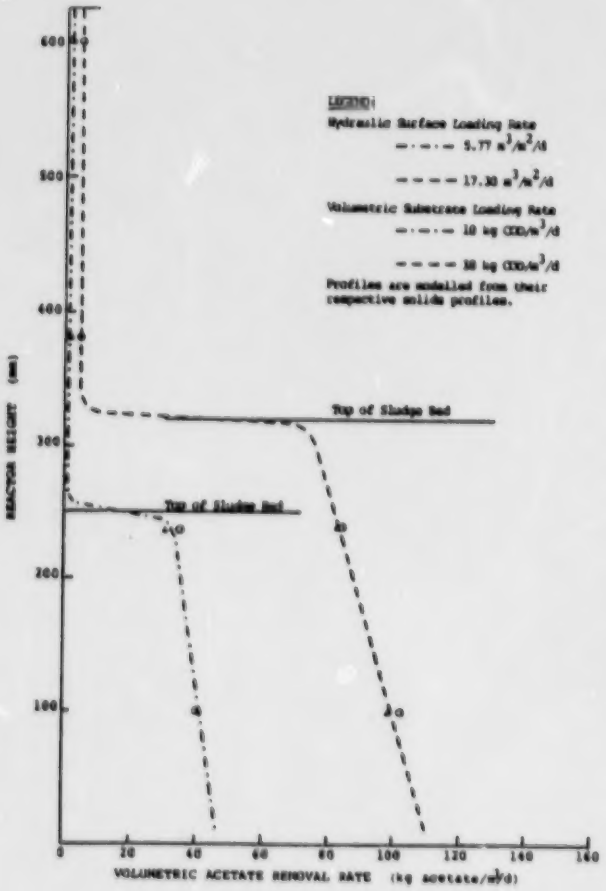


FIGURE 6: VOLUMETRIC MICROBIAL ACTIVITY DISTRIBUTION PROFILE

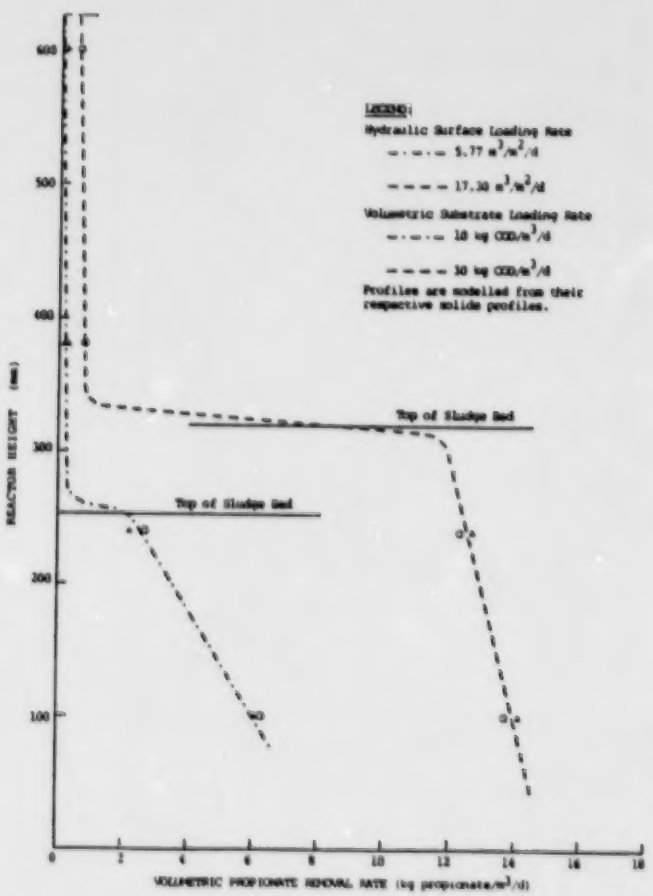


FIGURE 7: VOLUMETRIC MICROBIAL ACTIVITY DISTRIBUTION PROFILE

proceeding at its maximum potential rate. Further investigation of this matter may be required prior to scale up in order to optimize the dimensions of the reactor, the amount of sludge that can be put into a USB reactor (e.g. the depth of sludge bed) and the corresponding design of the feed inlet system.

Two - Stage Treatment System

The results of the research described in this paper indicate that the USB reactor may be ideally suited as the methanogenic reactor in a two-stage anaerobic wastewater treatment system. Such an application will result in the development and maintenance of stable granular highly-active sludge which in turn, will result in stable operating conditions and excellent treatment efficiencies at high loading rates.

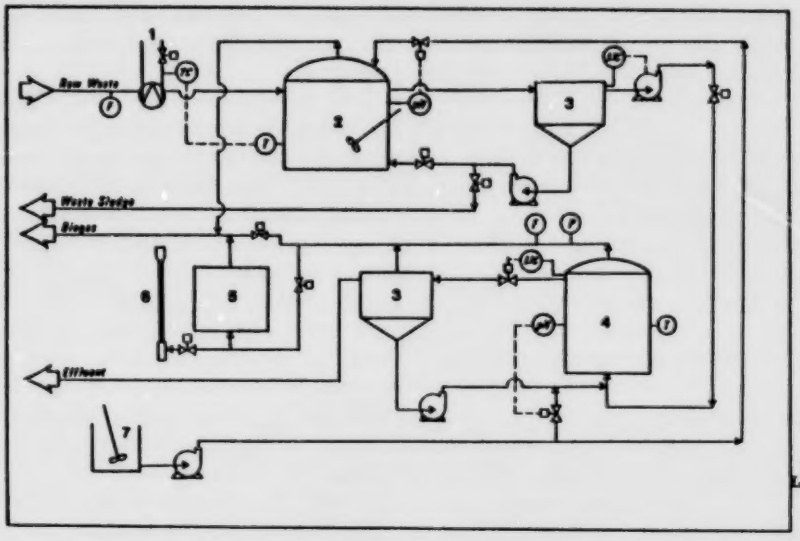
Both the design and operation of a two-stage anaerobic wastewater treatment system are functions of a number of factors:

1. The characteristics of the wastewater, including,
   - its temperature, and variations in temperature
   - the nature and concentration of the wastewater
   - variations in the composition and the COD of the wastewater
   - the pH and bicarbonate alkalinity of the wastewater
2. The sophistication of the treatment plant with respect to automation, control, etc.
3. The use that will be made of the biogas
4. The future expansion of the production capacity of the industry, if applicable
5. Environmental restrictions on the discharge of effluents to receiving waters or into municipal sewer systems.

Considering the wide range of these factors, the directions provided will be general, consisting of recommendations concerning the selection of process components. Emphasis is placed on the treatment of easily-hydrolyzed wastewaters. Such wastewaters are generally soluble. They are primarily composed of carbohydrates and originate from industries processing agricultural raw materials such as the sugar industry, the starch processing industry, and the brewing and distilling industry.

Figure 8 illustrates the schematic of a proposed two-stage anaerobic biological wastewater treatment system incorporating the USB reactor. Major components in the system are identified. Acidification may be carried out in a continuously stirred tank reactor (e.g. standard rate digesters) located upstream of the methanogenic reactor. This reactor should be dimensioned such that its void volume may accommodate peak influent wastewater flourates yet maintain the optional hydraulic retention time required for complete substrate turnover. Values for such design parameters have been recently evaluated by Zooetemeyer (21).

Alternatively the acidification may be achieved by retaining the wastewater in a holding tank for a specified period of time. Depending upon the nature of the wastewater to be treated, this measure would allow for a certain degree of spontaneous acidification to occur prior to introducing the waste into the methanogenic reactor (13). Although such a design would eliminate the requirement for an acidification reactor, it would produce non-homogeneous conditions of temperature and pH, subsequently, poor or incomplete acidification may result and negatively influence the operation of the methane reactor.



Major Components

1. Heat Exchanger
2. Hydrolysis and Acidification Reactor or Holding Tank
3. Sludge Settling Tank (Optional)
4. USB Methane Reactor
5. Gas Compressor Building (Optional)
6. Gas Flare
7. Flash Mixing Tank for Buffer

T = Temperature
F = Flow Meter
TC = Temperature Control
P = Pressure Control
pH = pH Indicator
LIC = Liquid Level Indicator Controller

Figure 8 : Schematic Diagram of the Proposed Two-Stage Anaerobic Wastewater Treatment Plant

Provisions should be made for the retention washed out of the acidification reactor at short liquid retention times, in order to maximize the concentration of biomass in the reactor and to minimize the growth of acid-forming micro-organisms in the USB methanogenic reactor. It has been shown that growth of acidogenic bacteria in the methane reactor causes deterioration of the settling characteristics of the methanogenic sludge. Thus, the acidification of the raw wastewater prior to its introduction into the methanogenic reactor should be as complete as possible. Provisions should also be made for a solid settling tank to be located following the USB methanogenic reactor, in order to maximize the retention and return of biomass, particularly during the start-up of the reactor. Alternatively, the USB reactor may be designed with an internal solids settler similar to full-scale designs currently in use in the USA and the Netherlands (22).

As previously stated, it is recommended that the USB reactor be used as the methane forming reactor. In this study, substrate loading rates of up to 45 kg COD/m^3/d (based on reactor volume only and not including external settler volume) and hydraulic shock loads could be easily accommodated. As describer earlier, the full scale design of the USB reactor, specifically as it concerns the feed inlet system and the quantity of granular sludge to be maintained in the reactor requires further investigation. Furthermore, provisions should be made in the design of the USB reactor for the recirculation of effluent. Although it was shown that the operation of the USB reactor with effluent recycle does not significantly improve the treatment efficiency of the process, it may be required to assist in the neutralization of the influent wastewater.

Depending on the size of the treatment system, the economics of collecting, compressing and possibly desulfurizing the produced biogas, and the local supply and cost of energy will determine the viability of utilizing the

biogas as an energy source. As illustrated in Figure 8, the proposed system is designed with a gas compressor facility. A back-up supply of natural gas from a utility company is also recommended.

Regarding process operating conditions, the acid phase reactor should be operated so as to maintain the composition of the acidified wastewater reasonably constant and thust stabilize the process in the subsequent methane reactor. The concentration of wastewater discharged from the acid forming reactor is expected to fluctuate with the concentration of raw industrial wastewater. Such fluctuations will result in abrupt surges in the substrate loading rate assuming a constant wastewater flowrate. Since it is generally acknowledged that the methanogenic reactor is very sensitive to such abrupt surges, it is recommended that the concentration of wastewater emanating from the acidification reactor be monitored and that the hydraulic loading rate to the methane reactor be adjusted in order to maintain a relatively stable substrate loading rate to this reactor. As demonstrated in the experimental study, the methanogenic treatment process is capable of effectively treating a large range of wastewater strengths and of tolerating sudden variations in wastewater strength while operating at a constant loading rate.

Variations in the relative concentrations of the main volatile acids (i.e. acetic, propionic, and butyric acids) emanating from the acid forming reactor are not anticipated to present operational difficulties in the USB methanogenic reactor. In the present experimental study a five fold fluctuation in propionic and acetic acid concentrations did not produce any noticeable impact on the treatment process. Recognizing that propionic is the slowest of the volatile acids to degrade, followed by acetic acid (13), fluctuations in other acid components such as butyric are not expected to produce negative effects on the operating conditions.

Finally, in order to determine scale-up factors influencing the full-scale operation of the two-stage treatment plant, pilot-scale studies of the two-stage process are recommended prior to full scale design and construction of such a plant.

CONCLUSIONS

a) The USB treatment process demonstrated the ability to effectively remove in excess 90 per cent of both the total COD and VFA-COD originally present in the raw wastewater, for waste strengths of 2 to 32g COD/L (1.7-27.2g VFA-COD/L) and hydraulic retention times varying between 3.2 and 76.8 hours, while operating at specific loading rates of 1.1 and 2.4 kg COD/kg VSS/d (0.9 and 2.1 kg VFA-COD/kg VSS/d). The treatment process demonstrated the ability to "polish" the effluent waste to VFA-COD concentrations of less than 300 mg/L. This is a very important consideration for an industrial waste producer located in a municipal setting, in that the requirement for further aerobic biological treatment of the waste would be minimal.

b) The microbial activity in the sludge bed of the USB reactor was observed to increase with an increase in the substrate loading rate applied to the treatment process. Acetate removal rates in the sludge bed were found to increase from a range of 1.3-2.3 kg acetate/kg VSS/d to 2.3-2.8 kg acetate/kg VSS/d, corresponding to an increase in the substrate loading rates from approximately 0.9 to 2.0 kg acetate/kg VSS/d, respectively. Similarly, propionate removal rates in the sludge bed were found to increase from approximately 0.2 kg propionate/kg VSS/d to 0.4 kg propionate/kg VSS/d.

c) The recirculation of effluent did not influence the treatment efficiency. Volatile fatty acid removal efficiency was maintained above 97 per cent for recycled effluent to feed ratios varying from 12 to 2. The necessity for the recirculation of effluent appears to be dictated by the pH of the influent waste stream. The recirculation of a portion of the effluent provides for a suitable pH in the influent zone by diluting the raw wastewater and adding biologically produced alkalinity, thereby neutralizing in part the influent waste stream.

d) The development of stable highly-active sludge granules, the ability to tolerate sudden variations in wastewater strength and the excellent removal efficiencies observed under different operating conditions when treating acidic soluble organic wastewaters demonstrate the feasibility of utilizing the USB reactor as the methane forming reactor in a two-stage anaerobic wastewater treatment process.

ACKNOWLEDGEMENTS

Financial assistance and technical support for this research was provided by the National Research Council of Canada and the University of Western Ontario. The co-operation extended by professional and support staff at both institutions is appreciated. Additional technical direction was provided by the Wastewater Technology Centre, Environment Canada.

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ANAEROBIC/AEROBIC TREATMENT OF MUNICIPAL LANDFILL LEACHATE

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ABSTRACT

Laboratory-model treatability studies were conducted to develop design guidelines and optimize treatment of high-strength leachate (23,000 mg COD/L, 17,500 mg BOD/L, 8,000 mg TOC/L, 1,000 mg Fe/L and 80 mg Zn/L) generated from a landfill site that serves the metropolitan area of Halifax, Canada. The selected processes included two treatment systems, I and II. System I consisted of an anaerobic fixed film reactor ($AFFR_1$), the effluent from which was further treated by an aerated lagoon and a settling lagoon in parallel. System II consisted of a second AFFR, deeper than the first. It was not followed by any other treatment.

The AFFR's were upflow cylindrical tanks with plastic "Tellerette" rings or loops which provided the fixed surface. The laboratory-model study at Halifax is the first known investigation which utilized the rings (patented by Pennwalt Corporation) that are manufactured as packing material in air pollution control devices.

The pH-adjusted, settled and nutrient-supplemented leachate (pre-treatment step) was added to $AFFR_1$, and $AFFR_2$ to maintain organic loads of about 2 and 1.6 kg COD/day.m^3, respectively. The two fixed film reactors were kept in a walk-in incubator (at 32 $\pm$ 1 °C) but the two lagoons were at room temperature (20 $\pm$ 2 °C).

The 12-week study showed that system Ia, consisting of pre-treatment, $AFFR_1$, and aerated lagoon, resulted in the best performance; system Ib, (pre-treatment, $AFFR_1$ and settling lagoon) generated the second best effluent, followed by System II, which produced an effluent of the highest organic and inorganic constituents.

System Ia resulted in the following removals: 99.4% COD, 99.9% BOD, 99.3% TOC, 99.9% Fe and 99.8% Zn. The corresponding values for System Ib were: 98.6% COD, 99.9% BOD, 98.5% TOC, 99.8% Fe and 99.8% Zn. System II (pre-treatment and $AFFR_2$), however, could remove only 97.3% COD, 99.1% BOD, 97.9% TOC, 99.1% Fe and 99.3% Zn. Similar data for the remaining 36 parameters are also included. Biogas generated by the two AFFR's has been estimated to range from about 0.4 to 0.57 m^3 per kg of COD destroyed.

An Engineer designing a treatment system can choose either System Ia or System Ib; though the latter produced an inferior effluent, it is an energy-efficient system and hence it should not be ruled out. Comparisons have been made between the performance potentials, capital and operating costs and energy producing/consuming potentials, of the Systems Ia, Ib and II.

INTRODUCTION

A municipal (sanitary) landfill site which is owned and operated by the Metropolitan Authority of Halifax, Dartmouth, Bedford and Halifax County (Metro Authority) receives about 700 tonnes of solid wastes per working day from the metropolitan area of Halifax (Porter Dillon Ltd., July 1984). Leachate generated at this site is collected in an underdrain system and is currently treated by two overloaded anaerobic lagoons in series which discharge to four siltation ponds, also in series, and then indirectly to the Sackville River via a large swamp. This overloaded system is rapidly reaching a critic1 point where reliability is questionable (Porter, 1982a and 1982b).

The Metro Authority retained the services of the consulting firm of Porter Dillon Ltd. to conduct a leachate treatability study in the summer of 1983. Laboratory-model studies were conducted at the Centre for Water Resources Studies (CWRS) of Technical University of Nova Scotia (TUNS), Halifax, to develop design guidelines for a proposed new treatment system which would enhance the currently used methods of treatment of the leachate. Field work, project co-ordination, cost estimates, and final design recommendations were completed by Porter Dillon Ltd.

BACKGROUND AND LITERATURE REVIEW

Grab samples of leachate were collected at the site and analyzed for about forty parameters. The results, summarized in Table 1, indicated that the raw leachate was very strong and, hence, a decision was made to explore the feasibility of using anaerobic fixed film reactors (AFFR) for handling this high-strength leachate. While the AFFR played the central role, the leachate had to be pre-treated for three reasons given in the next paragraph and post-treated to improve the effluent quality.

The first reason for pre-treatment is nutritional deficiency. According to Chian and DeWalle (1977), anaerobic filter treatment of leachate required a minimum P:COD ratio of 1:4360 and N:COD of 1:39. A review of data in Table 1 indicates that the raw leachate is deficient in phosphorus, but not in nitrogen. Therefore, the raw waste had to be phosphorus-supplemented. The second reason involves the potential for heavy metal toxicity. Though the early study by Chian and DeWalle (1977) indicated that heavy metals with maximum Cu and Zn concentrations of 0.9 and 0.2 mg/L, respectively, may have inhibited the anaerobic fermentation process, a latter study by DeWalle, Chian and Brush (1979) showed that the maximum soluble (0.45-μm membrane filtered) Cu and Zn concentrations of 11 and 15 mg/L, respectively, were not inhibitory to anaerobic digestion. Moreover, Kirsch and Sykes (1971) reported that at a total Zn concentration of 100 mg/L and a soluble concentration of 1.5 mg/L, gas production was inhibited. None of the studies observed a toxicity due to iron. Therefore, it was concluded that Zn was the only heavy metal which was potentially toxic to anaerobic microbes. Therefore, the raw leachate had to be pre-treated to precipitate Zn (by physical-chemical treatment). The third reason focuses on the low pH of the leachate; it is well documented that maintenance of stable pH values near the neutral range is necessary for successful anaerobic treatment (Ferguson and co-workers, 1984). Upward pH adjustment of leachate (from 5.6 to 7.2) was essential to meet this requirement.

Table 1 - Characteristics of Raw Leachate

Parameter	Mean Concentration*	Numbers of Samples Tested	Concentration Range
Alkalinity (as $CaCO_3$)	3,850.	12	3,450. - 4,580.
Aluminum	1.95	10	1.5 - 2.7
Ammonia (as N)	370.	12	285. - 406.
Antimony	0.18	10	<0.05 - <0.5
Arsenic	0.03	10	0.009 - <0.05
Barium	0.2	10	0.16 - 0.23
Beryllium	<0.005	10	<0.005 - <0.050
BOD	16,118	12	12,400. - 17,800.
Boron	6.5	10	5.1 - 8.9
Cadmium	0.004	10	<0.002 - 0.007
Calcium	1,741.	9	1,584. - 1,901.
Chloride	1,108.	9	930. - 1,320.
Chromium	0.2	10	0.10 - 0.28
COD	22,845.	12	19,900. - 24,100.
Cobalt	0.40	10	<0.010 - 1.78
Copper	0.03	10	<0.010 - 0.060
Hardness (as $CaCO_3$)	5,423.	12	4,873. - 5,916.
Humic Acid	416.	10	340. - 500.
Iron	937.	10	767. - 1,088.
Kjeldahl Nitrogen	490.	10	410. - 762.
Lead	0.10	10	<0.002 - 0.33
Magnesium	257.	8	223. - 284.
Manganese	58.9	10	47.5 - 68.1
Nickle	0.48	10	0.25 - 0.75
Nitrate + Nitrite (as N)	<0.5	12	<0.5 - <0.5
Orthophosphate	0.5	12	0.05 - <2.0
pH	5.6	12	5.4 - 5.7
Potassium	420.7	9	370. - 470.
Selenium	0.29	10	<0.1 - <1.0
Sodium	916.7	9	750. - 1,010.
Suspended Solids	467.6	12	110. - 1,150.
Sulphate	831.	12	750. - 930.
Tin	<0.03	10	<0.03 - <0.30
Total Dissolved Solids	15,297.	12	13,342. - 16,840.
Total Organic Carbon	8,136.	12	7,000 - 9,900.
Total Phosphate	9.1	10	6.6 - 17.0
Total Solids	15,734.	12	13,879. - 17,286.
Total Volatile Acids	10,100.	2	9,700. - 10,500.
Total Volatile Solids	6,187.	2	6,030. - 6,344.
Vanadium	0.25	10	0.16 - 0.54
Zinc	68.3	10	57. - 80.

*All concentrations, except pH, are expressed in mg/L
(Composition of the raw leachate samples used in this study is shown in Table 3)

EXPERIMENTAL PROCEDURE

Raw leachate was collected in a large quantity from a manhole in the landfill leachate collection system and stored in a freezer. Feedstock was prepared weekly from a thawed liquid of raw leachate. A local industry discharges lime slurry as a waste by-product onto the landfill site. Though the composition of the slurry varies from time to time, the range of concentrations (Tables 2A and 2B) of various constituents shows that it can be a very useful material for pre-treatment of raw leachate. Such an arrangement results in cost savings and encourages the concept of waste recycle.

Table 2A - Composition of Lime Slurry

Parameter	Range
slurry pH	12.1 to 12.7
pH of Supernatant (after 30-minute settling)	12.0 to 12.5
Total Solids	40,000 to 70,000 mg/L
Suspended Solids	20,000 to 60,000 mg/L

Table 2B - Chemical Composition of Dry Slurry (by weight)

Calcium	50.6%
Carbonate	4.1%
Hydroxyl	35.4%
Aluminum	6470 - 7020 ppm
Silica	6000 - 19,766 ppm
Iron	802 - 1,210 ppm
Magnesium	363 - 613 ppm
Manganese	13 - 97 ppm
Chromium	12 - 46 ppm
Lead	10 - 49 ppm
Copper	13 - 25 ppm
Zinc	5 - 10 ppm
Cadmium	0.05 - 5 ppm

(Source: Canada Liquid Air Company - unpublished report)

Table 3 - Typical Composition of the Feed and Raw Leachates Used in This Study

Parameter	Concentrations (mg/L) Raw Leachate	Feed Leachate
Alkalinity (as $CaCO_3$)	3,630.	4,200.
Aluminum	1.9	0.2
Ammonia N	375.0	380.0
Antimony	0.14	0.07
Arsenic	0.03	0.005
Barium	0.18	0.16
Beryllium	<0.005	<0.005
BOD	17,500.	17,400.
Cadmium	0.004	<0.002
Calcium	1,776.	2,025.
Chloride	1,060	---
COD	22,900.	22,300.
Cobalt	0.39	0.25
Copper	0.02	0.02
Hardness (as $CaCO_3$)	5,534.	6,044.
Humic Acid	445.	405.
Iron	1,002.	155.
Kjeldahl N	490.	470.
Magnesium	257.	---
Manganese	63.2	---
Nitrate + Nitrite (as N)	<0.2	<0.2
Orthophosphate (as PO_4)	0.2	1.0
Potassium	423.	---
Sodium	910.	---
Sulfate	930.	---
Suspended Solids	320.	277.
Total Dissolved Solids	14,940.	19,200.
Total Organic Carbon	8,100.	8,900.
Total Phosphate	17.	60.
Total Volatile Acids	10,100.	10,200.
Zinc	80.	4.

Pre-treatment. Jar tests were conducted to estimate the optimum pH range in which Zn can be insolubilized, to determine an optimum settling time for precipitation of Zn and estimate the volume of sludge produced. The results indicated that lime slurry addition to reach a pH within a range of 7.5 - 8.0, flocculation for 10 to 15 minutes at 30 to 40 rpm, and settling for one hour resulted in Zn concentration of 2 to 5 mg/L in the supernatant. The settled sludge volume was about 10 to 20 percent of the original leachate volume.

The supernatant pH was always lower than the pH value at the time of flocculation by about 0.2 to 0.5 unit. It further decreased, when stored in a refrigerator (at 4 $\pm$ 1 °C), by another 0.1 to 0.5 unit depending on length of storage.

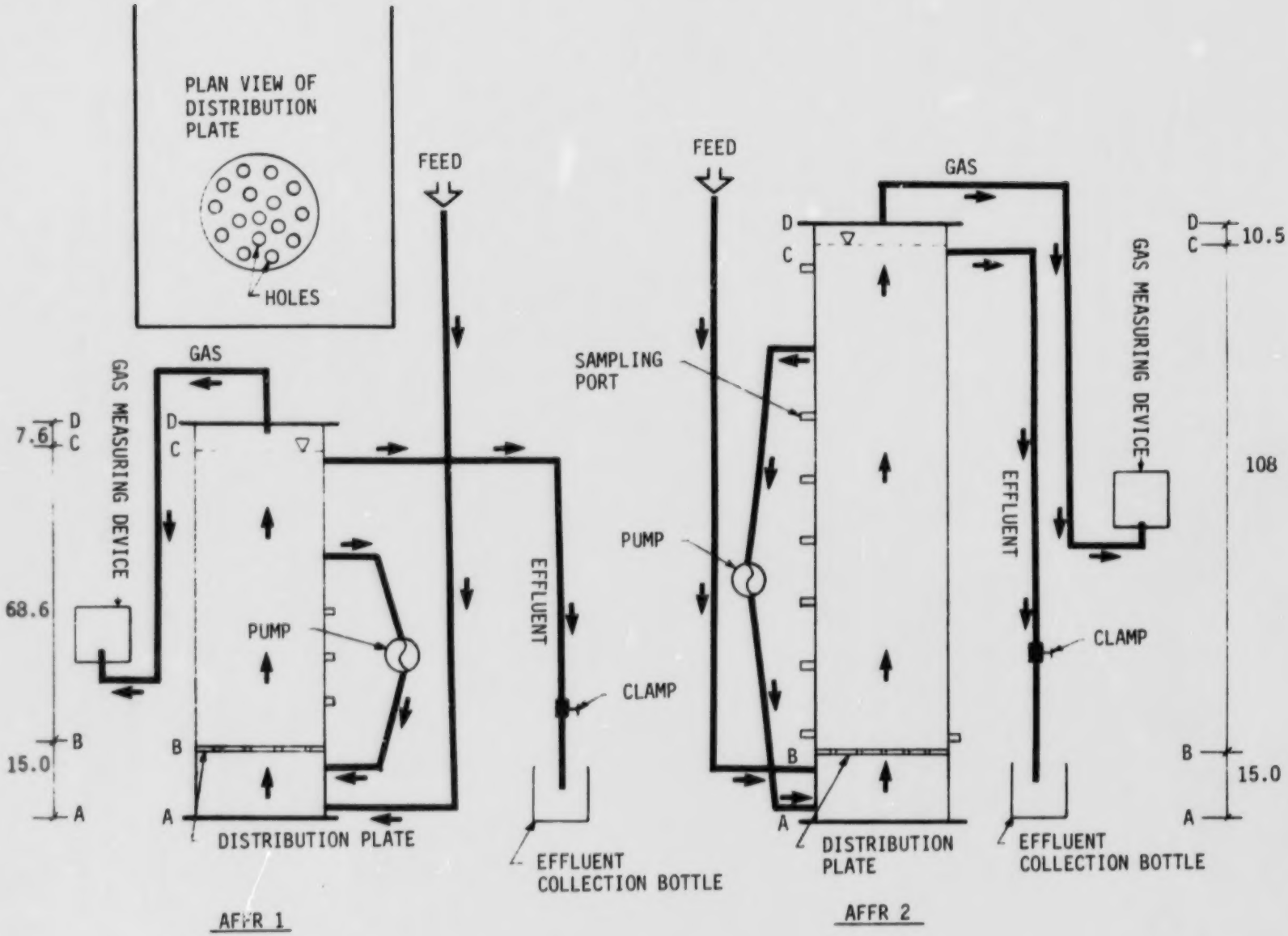
Based on jar test results, the following procedure was employed to pre-treat the leachate (and the resultant solution was labelled as feed leachate): lime slurry addition to increase pH to 8.0, settling the leachate for one hour, and drawing off of the supernatant (about top 85% of total volume). The supernatant pH, later on, was adjusted to 7.5 (if needed) and then a pre-determined quantity of Na_3PO_4 was added (just before feeding the anaerobic filters) to maintain a COD:P ratio of 1000:1. The phosphorus supplement is about 4.3 times the minimum recommended by Chian and DeWalle (1977). Typical composition of the feed leachate and raw leachate samples used in this study are shown in Table 3.

Anaerobic Fixed Film Reactors. The media in two laboratory-model reactors, AFFR1 and AFFR2, shown in Figure 1, consisted of randomly packed Ceilcote tellerette (registered trademark of Pennwalt Corporation) loops, the shapes of which are described by the manufacturer as 'filamentous toroidal helix'. Because of the filamentous media structure, there is less likelihood of dead space and short circuiting in the reactor. Table 4 summarizes the dimensions of the medium. Though the manufacturer designed them for use in air pollution

Table 4 - Physical Properties of the Medium

Parameter	Size
maximum dimension (outside diameter)	46 mm (1.81 inch)
loop height	19 mm (0.75 inch)
free volume	87%
solid volume	13% = 2,833 cc
surface area	180 m^2/m^3 (55 ft^2/ft^3)

control devices (scrubbers, stripping columns, etc.), this is the first known investigation in which this loop was used as the medium in an anaerobic fixed film reactor. The dimensions of the AFFR's are summarized in Table 5.



LABORATORY - MODELS OF ANAEROBIC FIXED FILM REACTORS
N.T.S. (ALL MEASUREMENTS IN CENTIMETERS)

FIGURE 1

Table 5 - Dimensions of the Two Laboratory-Model Filters

Unit	Inside dimensions (cm)				Inside Volumes (L)		
	Dia	Total ht AD*	Liquid ht AC*	Effective ht of medium BC	Distribution box volume AB	Empty Volume of Reactor BC (before adding the media)	Approximate Solid Volume of Media (TELLERETTE Loops)
AFFR 1	14	91.5	83.9	68.6	2.35	10.5	0.96
AFFR 2	14	134.0	123.5	108.0	2.35	17.0	1.53

* Points A, B, C and D are identified in Figure 1.

The media was supported on a 14cm diameter distribution plate. The feed leachate was added manually, twice a day, through the bottom opening just below the distribution plate which had 12 holes (of 12-mm diameter), uniformly placed (Figure 1); this enabled a uniform distribution of upflow through the reactor and minimized short circuiting. Recirculation in the upflow reactors was accomplished by using peristaltic pumps which recirculated reactor contents in AFFR1 at a rate of about 0.165 L/min and in AFFR2 at approximately 0.19 L/min. Recirculation pumps were operated between the hours of 9 am and 11 pm everyday. Measuring devices, to record the amount of biogas produced, were attached to the AFFRs, which were kept inside a walk-in room at 32 $\pm$ 1 $^{\circ}$C.

Aerated Lagoon. A glass aquarium tank of 21.1 L capacity and two diffuser stones supplied with compressed air acted as the model aerated lagoon (Figure 2). It was operated as a completely mixed aerated lagoon and the amount of compressed air injected was adjusted by eye judgement to provide good mixing in all parts of the aquarium tank. A portion of the effluent from AFFR1 was manually added (twice a day) to the inlet end of the aerated lagoon, which was kept at room temperature (20 $\pm$ 2 $^{\circ}$C). Pre-treatment, AFFR1 and this aerated lagoon represent system Ia.

Settling Lagoon. A tank identical to the model aerated lagoon was used as a settling lagoon; however, there was no aeration in this tank. A portion of the AFFR1 effluent was manually added (twice a day) to the inlet end of the settling lagoon. The performance of the aerated lagoon was compared to that of the settling lagoon, which was also at room temperature. Pre-treatment AFFR1 and the settling lagoon represent system Ib.

Pre-treatment and AFFR2 represent system II.

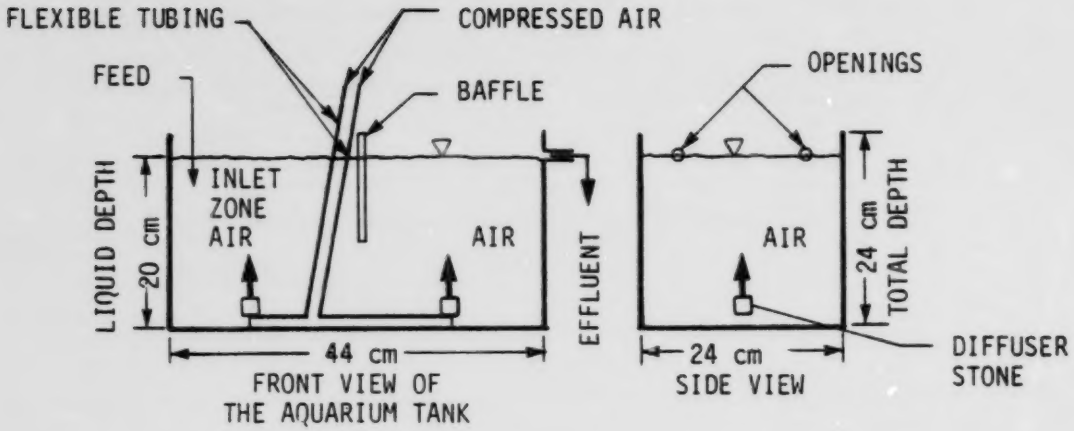


FIGURE 2 - LABORATORY - MODEL AERATED LAGOON

Start-up of Laboratory Models. The AFFR's were started with mixtures of equal parts of anaerobic municipal sludge from a local treatment plant's digester, contents of a laboratory-model anaerobic lagoon (from a previous study) and raw leachate. The initial contents were adjusted to a pH about 7.0 by using the waste lime slurry from a local industry. The initial incubation period was approximately four weeks.* The aerated lagoon was seeded with aerobically digested sludge from a local plant's aerobic digester. The initial content of the model lagoon consisted of a mixture of raw leachate, distilled water, AFFR1 effluent, and the aerobic sludge in the approximate ratio of 1:4:4:1. The mixture was brought to neutral pH and incubated for 20 days with aeration. The settling tank's initial content was a mixture of distilled water and AFFR1 effluent.

Laboratory Research Program

The two AFFRs and the lagoons were batch fed and operated for eight weeks. Including the four-week incubation period, the total duration of the investigation was 12 weeks. The two effluents from the AFFRs and those from the lagoons were tested for pH and temperature, twice a day. The four effluents were collected separately and the composite weekly samples were tested for TOC, COD, BOD, solids and nutrients every week; the tests for the remaining 20 parameters were conducted on three weekly composite samples during the eight-week study. The amount of methane produced by the two AFFRs was recorded twice a day. Loss due to evaporation from the four laboratory-model tanks was made up twice a day by adding distilled water. Daily feed rates, organic loading rates and hydraulic retention times for the four units are given in Table 6.

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* Prior to this study at 32°C, the reactors were used for a period of about 15 weeks at 20°C during which anaerobic microbes were acclimatized to the leachate.

Table 6 - Experimental Conditions and Loading Rates for the Four Laboratory - Models

Loading Rates	AFFR1	AFFR2	Aerated Lagoon	Settling Lagoon
Feed rate (L/day)	0.95	1.20	0.30	0.20
Hydraulic retention time (days)	12.5	14.9	70.3	105.5
Volumetric organic load ($Kg/m^3/day$)				
(8-week average)	2.0	1.6	0.013	0.009
(last 4-week average)	2.0	1.6	0.010	0.007

DISCUSSION OF RESULTS

The results are summarized in Tables 7 to 10 and Figures 3 to 6.

Table 7 gives the mean values of effluent parameters during the steady state period. AFFR1 reached a relative steady state during the last four weeks; however, the AFFR2 effluent quality stayed relatively uniform during most of the study period (Figures 3 to 6). Therefore the AFFR1 effluent values represent the last four-week averages and the AFFR2 effluent the whole eight-week averages. The term steady state is loosely defined herein; it represents a low degree of variations in the effluent quality, it does not indicate absolutely constant values. The aerated and settling lagoons reached steady state during the last four weeks. Some parameters, such as effluent pH, stayed exceptionally uniform, in all four units throughout the study period; in such cases the averages are based on eight-week values (the pH was measured twice a day; therefore 112 observations are indicated in Table 7). The other parameters which were reasonably uniform in concentration were suspended solids in AFFR1, AFFR2 and the settling lagoon, total volatile acids in the aerated lagoon and total dissolved solids in AFFR2 and the settling lagoon.

Removal efficiencies, during the steady state periods, are shown in Table 8.

Performances of Treatment Systems

Pre-treatment. Physical chemical treatment (lime precipitation at pH 8) resulted in high removals of Al (90%), As (83%), Fe (85%) and Zn (95%) and moderate removals of Sb (50%) and Cd (50%) and insignificant removals of Ba (11%) and Co (36%); however, organic matter, as measured by BOD, COD and TOC, could not be precipitated by lime (Table 3). Pre-treatment successfully removed the Zn which was potentially toxic and increased the pH, alkalinity (by 16%) and PO_4. Though the latter was added in the soluble ortho form (Na_3PO_4), most of it combined with lime to form insoluble complexes of calcium and phosphate; as shown in Table 3, orthophosphate increased moderately from 0.2 mg/L in the raw leachate to 1.0 mg/L in the feed

Table 7: Average concentrations of effluents during the steady state period

Parameter	Effluent concentrations in mg/L (except pH)			
	AFFR 1	AFFR 2	Aerated lagoon	Settling lagoon
Alkalinity (as $CaCO_3$)	2563(3)	--	812(2)	1800(2)
Aluminum	<0.5(3)	--	<0.5(3)	<0.5(3)
Ammonia (as N)	382(4)	369(8)	1.3(4)	37.5(4)
Antimony	<0.5(3)	--	<0.5(3)	<0.5(3)
Arsenic	<0.05(3)	--	<0.05(3)	<0.05(3)
Barium	<0.05(3)	--	<0.05(3)	<0.05(3)
Beryllium	<0.05(3)	--	<0.05(3)	<0.05(3)
BOD	254(4)	146(5)	4(4)	15(4)
Boron	6.1(3)	--	4.3(3)	6.2(3)
Cadmium	<0.02(3)	--	<0.02(3)	<0.02(3)
Calcium	108(2)	--	86(2)	27(2)
Chloride	1015(3)	--	790(2)	1080(3)
Chromium	<0.1(3)	--	<0.1(3)	<0.1(3)
COD	693(4)	625(8)	142(4)	325(4)
Cobalt	<0.1(3)	--	<0.1(3)	<0.1(3)
Copper	<0.1(3)	--	<0.1(3)	<0.1(3)
Hardness (as $CaCO_3$)	909(3)	--	707(2)	857(3)
Humic Acid	368(4)	--	230(2)	381(4)
Iron	11.8(4)	9.4(3)	0.36(4)	1.39(4)
Kjeldahl Nitrogen	395(4)	388(8)	6.5(4)	49(4)
Lead	<0.02(3)	--	<0.02(2)	<0.02(3)
Magnesium	154(2)	--	128(1)	187(2)
Manganese	0.6(3)	--	0.10(2)	0.07(3)
Nickle	<0.2(3)	--	<0.2(3)	<0.2(3)
Nitrate+Nitrite (as N)	<0.5(8)	<0.5(8)	138.0(4)	<0.5(8)
Orthophosphate	0.82(4)	0.9(4)	2.1(4)	5.6(4)
pH	7.43(112)	7.45(112)	8.56(112)	8.82(112)
Potassium	388(2)	--	299(2)	412(2)
Sodium	955(2)	--	775(2)	975(2)
Suspended Solids	117(8)	155(8)	4(4)	24(8)
Sulphate	28(2)	--	210(2)	54(3)
Tin	<0.3(3)	--	<0.3(3)	<0.3(3)
Total Dissolved Solids	4220(4)	4096(8)	3497(4)	4215(8)
Total Organic Carbon	224(4)	170(3)	59(4)	120(4)
Total Phosphate	5.2(4)	8.9(4)	3(4)	7.0(4)
Total Volatile Acids	55(4)	27(4)	20(8)	28(4)
Total Volatile Solids	714(4)	--	414(4)	611(4)
Vanadium	<0.1(3)	--	<0.1(3)	<0.1(3)
Volatile Suspended Solids	86(1)	73(7)	3(1)	20(1)
Zinc	0.52(4)	0.54(4)	0.13(4)	0.11(4)

The numbers in parentheses represent the number of observations.

Table 8 - Performance levels of the three systems and four treatment units (expressed as percent removal)

Parameter	System I(a)	System I(b)	System II	AFFR 1	AFFR 2	Aerated lagoon	Settling lagoon
Alkalinity (as $CaCO_3$)	77.6	50.4	--	39.0	--	68.3	30.0
Ammonia (as N)	99.7	90.0	1.6	(0.5)*	2.9	99.7	90.2
BOD	99.9	99.9	99.1	98.5	99.2	98.4	94.1
Calcium	96.2	98.5	--	94.7	--	20.4	75.0
Chloride	25.5	2.8	--	--	--	22.2	(1.5)
COD	99.4	98.6	17.3	96.9	97.2	79.5	53.1
Hardness (as $CaCO_3$)	87.2	84.5	--	85.0	--	22.2	5.7
Humic Acid	48.3	14.4	--	9.1	--	37.5	(3.5)
Iron	99.6	99.8	99.1	92.4	93.9	96.9	88.2
Kjeldahl Nitrogen	98.7	90.0	20.8	16.0	17.4	98.4	87.6
Magnesium	50.2	27.2	--	--	--	16.9	(21.4)
Manganese	99.8	99.9	--	--	--	83.3	88.3
Nitrate + Nitrite (as N)	**	**	**	**	**	**	**
Orthophosphate	***	***	***	18.0	10.0	(156.1)	(583)
Potassium	29.3	2.6	--	--	--	22.9	(6.2)
Sodium	14.8	(7.1)	--	--	--	18.8	(2.1)
Suspended Solids	98.8	92.5	51.6	57.8	44.0	96.6	79.5
Sulphate	77.4	94.2	--	--	--	(650.0)	(92.9)
Total Dissolved Solids	76.6	71.8	72.6	78.0	78.7	17.1	0.1
Total Organic Carbon	99.3	98.5	97.9	97.5	98.1	73.7	46.4
Total Phosphate	***	***	***	91.3	85.2	42.3	(34.6)
Total Volatile Acids	99.8	99.7	--	99.5	--	63.6	49.1
Total Volatile Solids	--	--	--	--	--	42.0	14.4
Volatile Suspended Solids	--	--	--	--	--	96.5	76.7
Zinc	99.8	99.8	99.3	87.0	86.5	75.0	78.8

* Values in parentheses indicate the percent increases.

** The concentrations being below detectable limits, % removals could not be calculated for these parameters and for Al, Sb, As, Ba, Be, Cd, Cr, Co, Cu, Pb, Ni, Se, Sn, and V.

*** These values were not calculated because phosphate was added as a pre-treatment step.

-- Not monitored.

Table 9 - Effluent pH Values

Week No.	AFFR 1		AFFR 2		Aerated Lagoon		Settling Lagoon	
	Range	Mean	Range	Mean	Range	Mean	Range	Mean
1	7.30-7.80	7.42	7.45-7.75	7.61	8.35-8.60	8.48	8.60-8.72	8.66
2	7.30-7.50	7.40	7.25-7.50	7.42	8.40-8.55	8.50	8.65-8.85	8.75
3	7.30-7.65	7.45	7.35-7.60	7.46	8.40-8.80	8.62	8.80-8.90	8.83
4	7.30-7.85	7.36	7.25-7.50	7.39	8.50-8.70	8.54	8.70-8.85	8.81
5	7.20-7.45	7.44	7.30-7.45	7.42	8.50-8.75	8.60	8.80-8.90	8.83
6	7.40-7.50	7.45	7.40-7.50	7.44	8.50-8.70	8.59	8.80-8.90	8.87
7	7.40-7.50	7.43	7.40-7.50	7.44	8.50-8.65	8.60	8.80-8.90	8.87
8	7.40-7.50	7.45	7.40-7.50	7.44	8.30-8.70	8.57	8.75-9.00	8.89
1 to 8	7.20-7.85	7.43	7.25-7.75	7.45	8.30-8.80	8.56	8.60-9.00	8.82

Table 10 - Estimated values of the costs and energy consuming/producing potentials of the three systems (for a leachate flow rate of one L/sec or 86.4 m^3/day)

Parameter	System I a	System I b	System II
Estimated capital cost (millions of Canadian dollars, 1984 values)	$1.76	$1.70	$1.46
Estimated operating cost per year (1984 values)	$130,000	$110,000	$100,000
Net energy comsumption/ production potential	26 KW (35 HP)	11 KW (15 HP)	11 KW (15 HP)

FIG. 3 EFFLUENT COD DATA

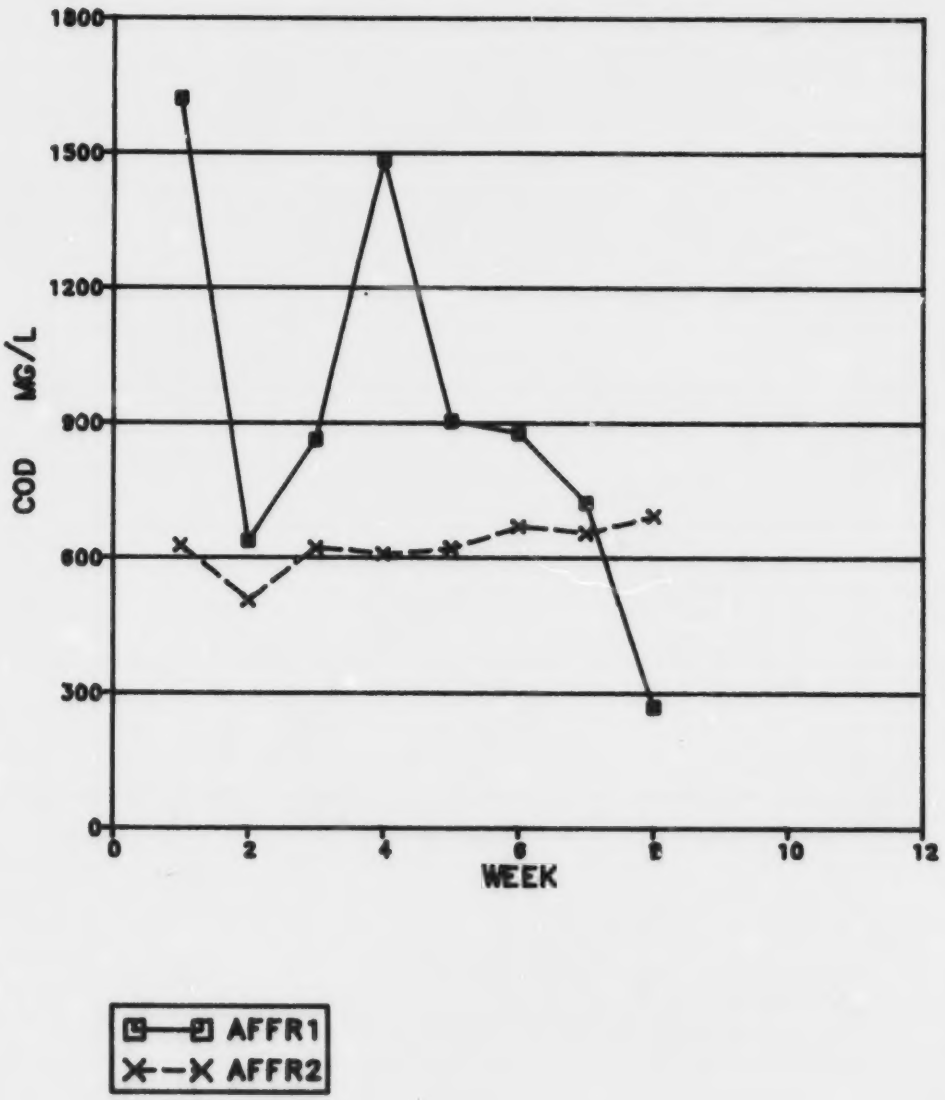


FIG. 4 EFFLUENT BOD DATA

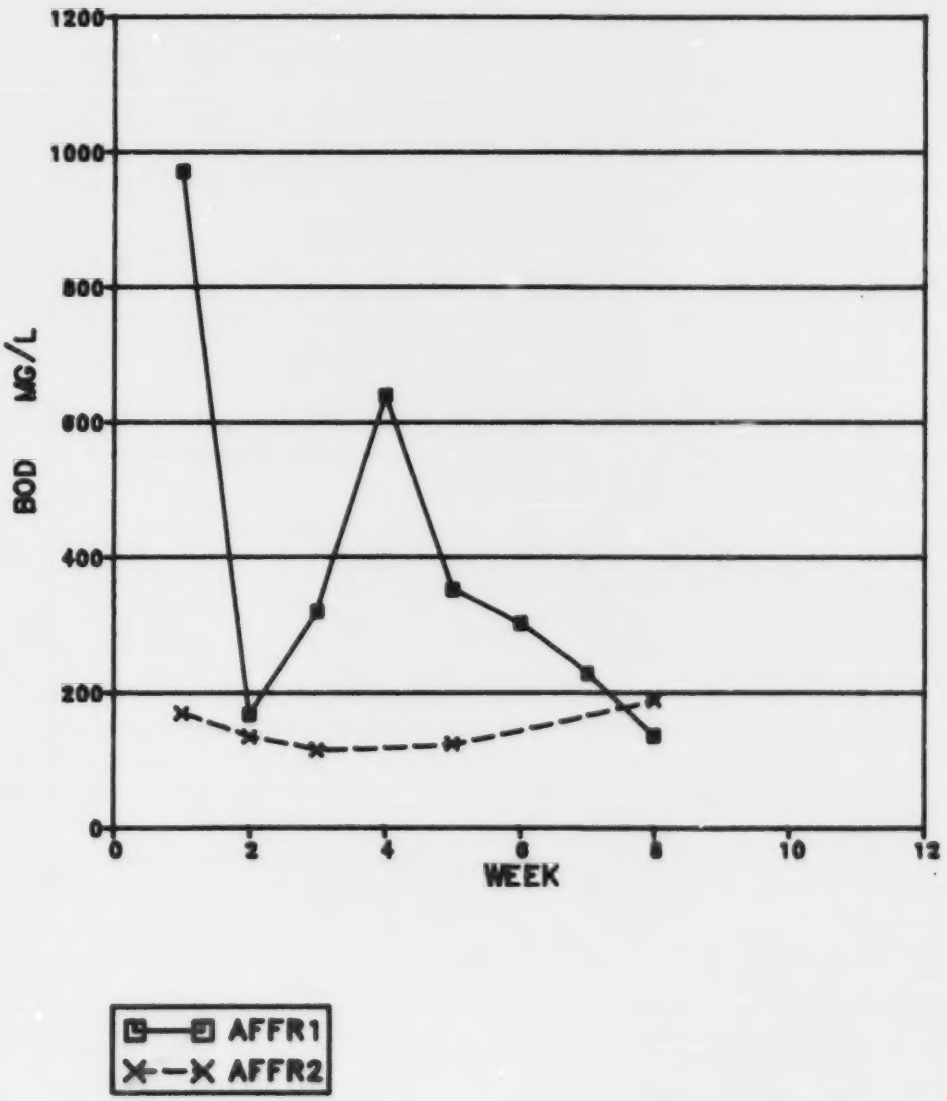


FIG. 5 EFFLUENT TOC DATA

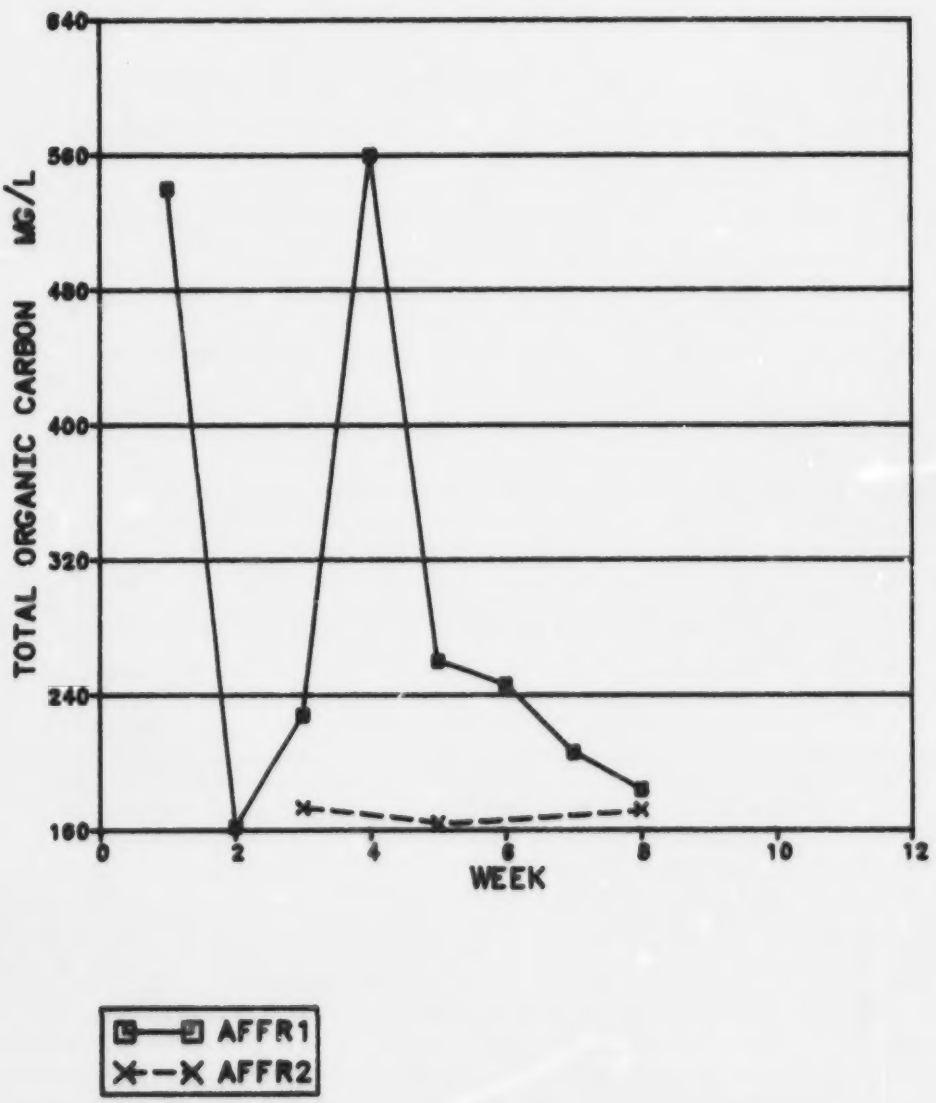
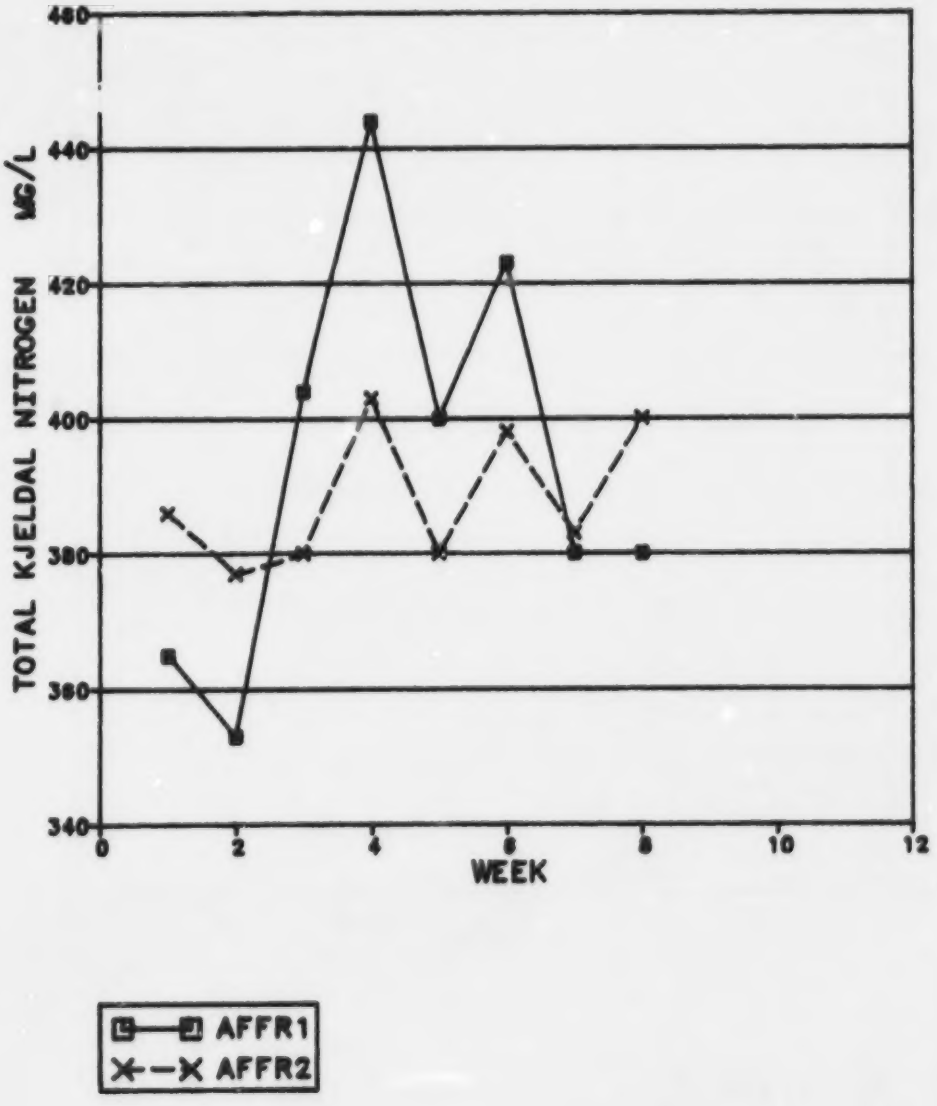


FIG. 6 EFFLUENT TKN DATA



leachate. However, total phosphate increased significantly from 17.0 to 60.0 mg/L. (Further studies are planned to focus on the specific phosphate requirements (type of PO_4 - ortho, condensed or organic --) for anaerobic treatment).

Anaerobic fixed film reactors. AFFR2 which was subject to a lower volumetric organic load (Table 6) but deeper than AFFR1 (Table 5), as expected, produced a better effluent with lower concentrations of BOD, COD, KN and TOC; however, AFFR1 produced an effluent of lower suspended solids (SS), total PO_4 and Zn (Tables 7 and 8). The superior performance of AFFR2 (in the removal of organic matter) is due to the lower volumetric organic load; but the inferior performance from the viewpoints of SS, total PO_4 and Zn is possibly due to higher rates of mixing. A higher capacity pump used for mixing the contents of AFFR2 could have resulted in keeping a larger proportion of the (insoluble) solids in suspension in that filter; whereas a smaller pump used for AFFR1 could have resulted in lower mixing levels and the subsequent settling of higher proportion of SS, total PO_4 and Zn. (About 75% of the Zn was in the suspended form and the balance in the dissolved form).

Though the normal range of volumetric organic loads used in several studies is from 2 to 8 kg COD/day/m^3 of AFFR, a wide range of 0.5 to 40 kg COD/day/m^3 has been quoted in the literature (Hall, et al, 1982; Henze and Harremoës, 1983; Van den Berg and Kennedy, 1982 and Young, 1983). In comparison, the loading rates (of 2 and 1.6 kg COD/day/m^3) used in the current study were rather in the lower range because an attempt was made to produce a good quality effluent. A higher loading of say, 3 to 5 kg/d/m^3 would produce an effluent of inferior quality and it could tax the downstream aerated lagoon resulting in higher energy cost. (However, a pilot scale study of a downflow filter--developed by National Research Council of Canada--has just been completed by the senior author of this paper at loading rates varying from about 1 to 5 kg COD/d/m^3). Based on this pilot scale study and the current laboratory model study, a loading rate of 3 kg COD/d/m^3 has been recommended to the Halifax Metro Authority (Porter Dillon Ltd. July 1984).

The amount of biogas produced by AFFR1 varied from 0.39 to 0.54 m^3 per kg of COD destroyed and by AFFR2 from 0.43 to 0.54 m^3 per kg of COD destroyed. The arithmetic mean values of biogas produced were 0.48 and 0.52 m^3 per kg of COD removed by AFFR1 and AFFR2, respectively.

During the course of the current 12-week study, no practical problems such as clogging of filter medium were encountered. The Tellerette loops/rings seem to have good potential for use in upflow anaerobic fixed film reactors.

The orthophosphate test was conducted on the filtrate of all samples (raw leachate, feed leachate and the four effluents). The total phosphate test was, however, conducted on the full (unfiltered) samples. As can be seen from Table 3, most of the orthophosphate added (in the form of N_3PO_4) combined with lime and was recorded as total phosphate. It is well known that soluble orthophosphate (also called soluble reactive phosphorus) is much more readily available for microbial use than the remaining portions of the total phosphate. Therefore, the phosphate (or phosphorus) requirements should, preferably, be expressed in two ways. Accordingly, it may be concluded that an orthophosphate:COD ratio of 1:22,300 and total phosphate:COD ratio of 60:22,300 (Table 3) are adequate for anaerobic metabolism. When expressed as phosphorus (P), ortho P:COD ratio of 1:7430 and total P:COD of 1:124 could sustain anaerobiosis. When comparing the P requirements of this study with

that of Chian and DeWalle (1977) (who suggested a minimum total P:COD ratio of 1:4360 in their study of leachate treatment), the following points of similarities and dissimilarities of the two investigations should be taken into account. They stated that, due to the initial seeding of the unit with digested sludge, "no nutrient additions were required during the 518 day period even though the COD:P and COD:N ratios were as high as 4360:1 and 39:1, respectively." Chian and DeWalle used sodium hydroxide for increasing their leachate pH from about 5.4 to 7.0, and the effluent had about three-fourths of the total phosphate in the suspended solids form. A comparison of the two studies shows that though both of them used digested sludge as the seed, three dissimilarities exist between them: two different chemicals were used for pH adjustment, one study included a continuous addition of P supplement but not the other, and P requirement is expressed as total P in one study but as ortho P as well as total P in the other. Therefore, precise comparisons cannot be made. Considerable amounts of gaps seem to exist in the literature in the area of phosphate biochemistry associated with anaerobic metabolism. Therefore additional studies are planned for the future to clarify some of the currently unknown aspects related to this issue.

The feed leachate had, typically, about 4 mg/L of Zn (Table 3), approximately 75% of which was attached to the suspended particles and the balance was in the dissolved form (these values varied, depending on how carefully the supernatant leachate was decanted, during the weekly batch pre-treatment operation in the laboratory). No evidence of Zn toxicity could be detected during the course of this study; removal of biodegradable organic matter exceeded 98% and considerable amount of biogas was generated. Therefore, it may be concluded that Zn, at a concentration of 4 mg/L, is not toxic to anaerobic microbes in a well mixed AFFR.

The two lagoons. The aerated lagoon, as expected, considerably outperformed the settling lagoon in several respects. With the exceptions of Ca, Mn, SO_4 and Zn, the aerated lagoon effluent concentrations were better than the settling lagoon effluent concentrations (Table 7). Two possible reasons could be suggested for the lower levels of Ca, Mn, SO_4 and Zn in the settling lagoon which was operating consistently at a higher pH level than the aerated tank (Table 9). Solubilities of most metals decrease when pH increases; therefore, Ca, Mn and Zn could have settled in the settling tank more effectively than in the aerated lagoon. The sludge mass settled at the bottoms of the two tanks were not tested for Ca, Mn and Zn at the end of the test; therefore this reasoning cannot be confirmed. Though sulfide and other reduced forms of sulfur were not monitored in this study, it is reasonable to assume that high sulfate concentration in the leachate was converted to the reduced forms in the anaerobic filter which were in turn biologically oxidized in the aerated lagoon back to sulfate. As a result, sulfate concentration increased from 28 mg/L to 210 mg/L in the aerated lagoon. Moreover, pronounced degrees of ammonification and nitrification took place in the aerated lagoon resulting in high removals of ammonia N and KN. Very high removals of BOD (98%), ammonia N (99%) and KN (98%) are due to long retention time (70 days); however, the removals of COD (79%) and TOC (73%) were lower, indicating the presence of nonbiodegradable organics in the aerated lagoon effluent.

System Ia. Pre-treatment, AFFR1 and aerated lagoon resulted in COD reduction from 22,900 mg/L to 142 mg/L, BOD from 17,500 mg/L to 4 mg/L and TOC from 8100 mg/L to 59 mg/L (Tables 3 and 7). Similarly, impressive metal removals of Zn (from 80 to 0.13 mg/L) and Fe (from 1002 to 0.36 mg/L) were accomplished. However, nitrification in the aerated lagoon resulted in a very high effluent nitrate concentration of 138 mg/L. The aerated lagoon effluent will be discharged through a series of existing "siltation" (settling) ponds in which denitrification could potentially take place. Further studies with shorter-term aeration followed by settling tank(s) would focus on this problem of excess nitrates. Nevertheless, effluent generated from the new system will be of much better quality than the effluent currently produced (Porter, 1982a and Porter Dillon Ltd., July 1984).

System Ib. Though the settling lagoon produced a generally inferior effluent than the aerated lagoon, it had one superior quality, namely, lower nitrate concentration of less than 0.5 mg/L, as N (Table 7). Assuming that the laboratory model settling lagoon simulates one or more of the existing "siltation ponds" in the field (there is no reason to assume otherwise), one of the two following options may be employed: operate the new aerated lagoon and siltation ponds in series or in parallel.

System 2. The effluent from AFFR2 was not analyzed as extensively as the other three effluences, due to budgetory constraints. It produced an effluent of quality better than AFFR1 but not as good as Systems Ia and Ib.

The Recommended Treatment System. A summary of the estimated capital costs and operating costs and the energy consumption/production potentials of systmes Ia, Ib and II are given in Table 10. The values represent preliminary estimates for a design flow rate of one L of leachate per sec (86.4 m^3/day). Taking into account the facts that system Ia produced the best effluent--except for excess nitrate concentration--and system Ib generated an effluent of low nitrate concentration, it is recommended that a combination of Ia and Ib be used. The effluent from AFFR can be treated by aerated and settling lagoons in series or parallel.

CONCLUSIONS

The following conclusions can be drawn from this study.

1. The Tellerette rings or loops (patented by Pennwalt corporation) originally designed for use in air pollution control devices has been successfully used, first time anywhere, as a filter medium in two laboratory-model anaerobic fixed film reactors (AFFRs). Additional studies with pilot scale models would encourage engineers to use the medium in prototype AFFRs.
2. At volumetric organic loads of 1.6 and 2 kg COD/day/m^3, 32 + 1°C and with an influent of pH of 7.5 and alkalinity of 4200 mg/L (as $CaCO_3$), the filters could stabilize between 96 and 99% of organic matter as indicated by BOD, COD and TOC removals.
3. Zn, at a concentration of 4 mg/L, does not seem to exhibit any toxicity to anaerobic microbes.

4. Because waste lime slurry was added to the leachate for pH adjustment, only a small percent of the originally added orthophosphate (Na_3PO_4) remained in the soluble ortho form; a vast majority of it was converted to the other forms and measured as total phosphate. Therefore the levels of phosphorus requirements are suggested in the following two ways: (1) total P:COD of 1:124 and (ii) orthophosphorus:COD of 1:7430. At these levels no deficiency was detected.

5. System Ia, consisting of pre-treatment, anaerobic fixed film reactor 1 (AFFR1) and aerated lagoon, removed 99.4% COD, 99.9% BOD, 99.3% TOC, 99.9% Fe and 99.8% Zn. Long hydraulic retention time (70 days) in the aerated lagoon, however, resulted in "excessive" nitrification leading to 138 mg/L of nitrate + nitrite (N) in the aerated lagoon effluent.

6. System Ib, consisting of pretreatment, AFFR1 and settling lagoon resulted in 98.6% COD, 99.9% BOD, 98.5% TOC, 99.8% Fe and 99.8% Zn. Nitrification was not a problem in the settling lagoon which, however, produced an effluent of 37.5 mg/L of ammonia N and 49 mg/L of Kjeldahl N.

7. Combining the advantages of the two types of lagoons, it may be suggested that both of them be used either in parallel or in series.

ACKNOWLEDGEMENTS

This investigation was primarily supported by a contract funded by the Metropolitan Authority of Halifax, and secondarily by an operating Research Grant from Natural Sciences and Engineering Research Council (NSERC) of Canada. The authors gratefully acknowledge the assistance of Mrs. D. Blackmore and Miss Paula Power who typed the manuscript.

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KINETICS OF AN ANAEROBIC-AEROBIC REMOVAL OF 4,6-DINITRO-O-CRESOL

by

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SUMMARY

The removal of 4,6-dinitro-o-cresol (DNOC), a phenolic priority pollutant, and the removal kinetics were investigated using an anaerobic fluidized bed reactor followed by an activated sludge reactor.

The DNOC was completely converted during the anaerobic pretreatment stage (anaerobic bed effluent DNOC concentration <<1 mg/l), when the influent DNOC concentrations were as high as 600 mg/l. It was also found that, while 100% conversion of DNOC occurred during the anaerobic pretreatment stage, there was approximately 25% COD removal. The subsequent aerobic activated sludge stage reduced the anaerobic stage effluent COD by approximately 80%, resulting in about 90% overall removal.

Batch test studies with the anaerobic microorganisms provided preliminary data that were used for establishing a range of DNOC loading rate onto the anaerobic fluidized bed. The batch test studies also indicated that DNOC was highly toxic to the anaerobic microorganisms in the absence of a readily biodegradable co-substrate, and could not be used as a sole carbon source by the anaerobic bacteria. Sucrose was used as the co-substrate in this investigation. The anaerobic DNOC biodegradation was found to be a function of sucrose concentration. The DNOC degradation kinetics were modeled.

Previous investigation of the aerobic treatment of DNOC using conventional activated sludge process showed that DNOC removal is less than 25% and the concenration of DNOC that is tolerated by activated sludge microorganisms is only about 50-60 mg/l. The results of the present investigation demonstrate that the anaerobic-aerobic treatment process is an effective process for the removal of DNOC.

INTRODUCTION

4,6-Dinitro-o-cresol (DNOC), one of the U.S. EPA priority pollutants, is a yellow crystalline compound derived from

ortho-cresol. The compound is produced by direct nitration of o-cresol with a nitrating mixture at a low temperature. In some cases the o-cresol is first sulfonated with 70-93% sulfuric acid. DNOC is obtained in small amounts also from the oxidative nitration of toluene in the presence of mercuric nitrate (11).

DNOC is used primarily as a blossom-thinning agent on fruit trees and as a fungicide, insecticide, and miticide on fruit trees during the dormant season (16). It also has limited use in the dyestuff industry and for other minor miscellaneous industrial purposes (13).

DNOC has been detected in the following industrial wastewaters: organics and plastics manufacturing industries, pesticides manufacturing industry, rubber manufacturing industry, wood preservatives industry, pharmaceutical manufacturing industry, coal mining industry, foundries industry, and iron and steel manufacturing industry (12,17).

DNOC has been shown to undergo biodegradation by pure soil microorganisms cultures under aerobic conditions (5,8,15). However bench scale activated sludge studies (4) show that the removal of DNOC in the activated sludge process is less than 25%, and the concentration of DNOC tolerated by activated sludge microorganisms is only about 50-60 mg/l. Additional data on the treatment of DNOC in activated sludge are reported for synthetic wastewater as 58% removal, and the average achievable concentration was 2.1 mg/l (17,14) confirmed the insignificant aerobic biodegradation of DNOC by static-culture-flask biodegradation tests.

There is a paucity of information about the anaerobic degradation of most organic compounds, but some data indicate that aromatic nitro groups can be reduced to amino groups under anaerobic conditions (2,6). Results from studies on the degradation of DNOC by ruminal microorganisms confirmed the anaerobic reduction of DNOC (3,7). It was also shown that some nitro-aromatic compounds that can be degraded under anaerobic conditions are not degraded under aerobic conditions (9,10).

The treatment of DNOC by a single anaerobic process alone bears the risk of inactivating the environment-sensitive methanogens, in which case the organic loading of the pollutant then will not be reduced. It is also known that the benzene nucleus, which will probably still exist if DNOC is cometabolized to another aromatic compound during the anaerobic process, will undergo ring cleavage in the presence of oxygenase under aerobic conditions. An anaerobic pretreatment-aerobic treatment scheme for treating wastes containing DNOC is thus a very promising method to biodegrade the DNOC, as well as lowering the organic loading of the waste.

The objectives of this study, therefore are:

- To provide data on the mechanism and rate of the degradation of DNOC under anaerobic conditions.
- To examine the feasibility of anaerobic-aerobic treatment of DNOC by pilot plant study.
- To monitor the changes in toxicity due to the anaerobic-aerobic treatment of DNOC.

EXPERIMENTAL

Anaerobic Batch Reactors

Two type of anaerobic batch reaction studies were performed: shaker bottle and continuously-stirred. Both were run to provide preliminary data on the biodegradability of DNOC under anaerobic conditions.

Anaerobic Shaker Bottle Batch Test Apparatus and Procedures. Screw capped glass 240 ml bottles were used. They were incubated in a temperature controlled shaker water bath for seven days.

Each batch run starting with the same initial DNOC concentration. Eight bottles were run simultaneously, one being removed at 24 hour intervals over the seven day period. Initially, the bottles were purged with oxygen-free nitrogen, and 170 ml of DNOC-feed solution was added to each. The feed solution was made of varied concentrations of sucrose, $NaHCO_3$, $(NH_4)_2HPO_4$ and trace minerals in dechlorinated tap water, with a ratio of COD:N=100:5. DNOC was added to the feed solution from an aqueous alkaline stock solution of 10,000 or 20,000 mg/l DNOC, to achieve the desired final DNOC concentration. The solution was then mixed with nitrogen. A sufficient amount of anaerobic column effluent was collected and after brief settling under anaerobic conditions, each bottle was inoculated with 30 ml of the settled anaerobic column effluent. Each bottle then contained a final volume of 200 ml. Immediately following the inoculation, each bottle was purged with nitrogen and sealed. One of the eight bottles was used as animitanl sample and its contents analyzed for pH, oxidation-reduction-potential (ORP), suspended solids (SS), volatile suspended solids (VSS). Samples for soluble chemical oxygen demand (COD), soluble total organic carbon (TOC), and total organic acids (TOA) were also prepared. A sample was extracted and then analyzed by gas chromatography (GC) for DNOC concentration. Since DNOC feed solution contained sucrose as a co-substrate, an additional sample was analyzed for residual sucrose concentration. The remaining seven bottles were incubated in the 36.5°C water bath, the shaker was adjusted to a rate of 100 strokes per minute. During the first few days of the incubation period

there was considerable gas production. It was, therefore, important to relieve the gas and pressure produced by slightly opening the bottles and resealing them daily.

Amber glass bottles with 4.2 liters capacity each fitted with a three-hole rubber stopper was used as a reactor vessel. Heating tapes were used to maintain the temperature in the bottles at 36.5°C ± 1°C. Each batch was continuously-stirred with a magnetic stirrer. Additional bottles equipped with a two containing gas displacement fluid (aqueous Na_2SO_4-H_2SO_4 solution prepared following Method 511A, Standard Methods (1) each equipped with a two-hole rubber stopper.

Each bottle was filled with 3.75 ℓ dilution water (tap water) minus the volume of DNOC stock solution. After the stirers were started, the batch temperature was equilibrated to 36.5°c ± 1°c. The batch bottles and the bottles containing gas displacement fluid were purged with oxygen-free nitrogen. Sucrose, $NaHCO_3$, $(NH_4)_2HPO_4$, trace minerals and DNOC stock solution were added to each batch test bottle to make the desired DNOC and sucrose concentrations with total volume of 4 liters. Anaerobic column effluent was collected and 250 ml of it were used as inoculum for each bottle. After inoculation, the top space of each batch test bottle was quickly purged with nitrogen and each stopped tightly to ensure anaerobic conditions. A volume of 250 ml was withdrawn as an initial sample, by opening a clamp and siphoning with a section of tubing filled with water. Thereafter, a sample of 250 ml was withdrawn every 24 hours for the following seven days. After daily samples were taken they were analyzed for pH, ORP, SS, VSS. Samples for COD, TOC and TOA were also prepared. A sample was extracted and then analyzed by GC for the determination of DNOC concentration. When necessary, an additional sample was analyzed for residual sucrose (co-substrate) concentration.

Anaerobic-Aerobic Pilot Scale. The pilot plant consisted of an anaerobic upflow sand filter and an activated sludge unit, as shown in Figure 1. The anaerobic upflow sand filter consisted of a plexiglass column 8 feet high. Its lower section was 5 feet high and 3 inch ID and its upper section was 3 feet high and 4.25 inch ID. Prewashed sand, which was used as matrix for the attachment of microorganisms, was 3.0 feet high in the lower section. When operating, the bed expanded about 30% to a height of 4.0 feet. The temperature of the column was maintained at 36.5°C by heating tapes.

The anaerobic upflow sand filter column was operated as a continuous flow system with a recycle flow rate maintained at about 3,000 l/day. Sodium chloride tracer tests showed that the liquid phase was completely mixed. However, the biomass in the fluidized bed was not uniformly distributed. The microorganisms were predominately concentrated in the upper 22" ± 2" section of the sand

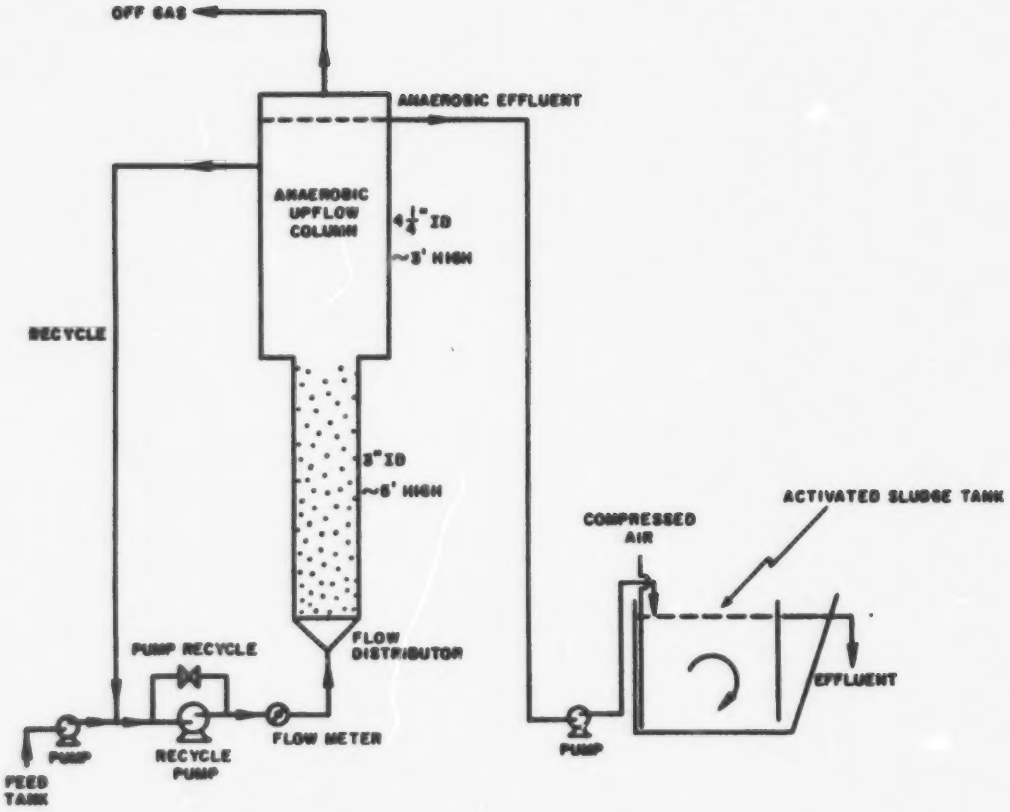

Figure 1. Schematic of pilot plant apparatus.

bed. The pattern of the biomass distribution in the column bed resulted in a distinct dividing plane formed between the sand particles in the lower section of the bed which were sparsely populated with biomass, and the sand particles in the upper section of the bed which were densely populated with biomass. The biomass distribution in the fluidized bed appears in Table 1.

TABLE 1. BIOMASS DISTRIBUTION IN THE FLUIDIZED BED

Sample Source	VSS (mg/l)
Anaerobic bed upper section	10,700 - 11,300
Anaerobic bed lower section	1,940 - 3,050
Anaerobic column effluent	12 - 104

The activated sludge tank was a conventional 19 liters plexiglass unit with a 14 liter aeration compartment and a 5 liter clarification compartment. Effluent from the anaerobic upflow column was pumped directly into the activated sludge unit as feed.

The anaerobic upflow sand filter column was started up by several inoculations with seed from a sludge digester of a local municipal sewage treatment plant. Operating parameters such as organic loadings, hydraulic retention times, loading concentrations of DNOC and of sucrose were carefully monitored.

The performance of the anaerobic column in the pretreatment process of DNOC was monitored by sampling its effluent and analyzing for pH, alkalinity, ORP, SS, and VSS. Samples for COD and TOC were also prepared. A sample was extracted and then analyzed by gas chromatography (GC) for determination of DNOC concentration. When necessary, an additional sample was analyzed for residual sucrose concentration. When gas production in the column was significant, the volume of the off-gas produced was analyzed and its volume was measured.

The performance of the activated sludge process in the treatment of the anaerobic-degradation-products was monitored by sampling its effluent and analyzing it for pH, COD and TOC.

Analytical Procedures

Experimental procedures employed in this study conformed to established laboratory techniques, and were in accordance with those described in "Standard Methods", "Methods for Chemical Analysis of

Water and Wastes", or "Test Methods: methods for organic analysis of municipal and industrial wastewater". Chemical oxygen demand (COD) was measured by the modified semi-micro tube method. The determination of DNOC concentration was done by gas chromatography (GC), using Varian Aerograph 1440 gas chromatograph with flame ionization detector. The column was a Supelco, Inc. 3 ft x 2 mm ID glass with 1% SP-1240 DA on 100/120 Supelcoport packing. The carrier gas was prepurified N_2 at 20 ml/min and 24 psig. The flame ionization detector fuel was prepurified H_2 (29 ml/min, 15 psig) and air (300 ml/min, 30 psig). Operating temperatures: 180°C injector and 180°C detector and the column temperature program was 2 min initial hold at 150°C, 15°C/min to 170°C and hold. The procedures for sample preparation, extraction, quality control and data interpretation followed Method 604-Test Method for Phenols (1).

Volumetric measurements of off-gas produced in the anaerobic column were done following volumetric method-511A of Standard Methods (1).

Off-gas composition was determined following gas chromatographic method-511B of Standard Methods (1), using sealed gas sampling bags for sample collection. Baseline Industries, Inc. Model 1030A gas chromatograph equipped with a thermal conductivity detector was used. Helium was used as carrier gas, analytical grade CO_2, CH_4 and N_2 were used as calibration gases.

The determination of oxidation-reduction potential (ORP) was done by the use of Sensorex reference combination ORP electrode (Sensorex Inst. Co., Westminister, California) along with a Fisher Accumet® 325 pH/Ion Meter (Fisher Scientific Co., Pittsburg, Pennsylvania). The pH was measured with a Fisher Accumet® 325 pH/Ion Meter (Fisher Scientific Co., Pittsburg, Pennsylvania).

The determination of sucrose concentration was done by using the indirect method in which sucrose is inverted by hydrolysis to give equal parts of glucose and fructose. Samples and sucrose standard curve samples were hydrolyzed with H_2SO_4 and the concentration of glucose formed after the inversion was determined by Beckman Glucose Analyzer II.

Suspended solids (SS) were determined using Method 209D of Standard Methods (1) for total nonfiltrable residue dried at 103-105°C, and volatile suspended solids (VSS) were determined using Method 209G for volatile and fixed matter in nonfiltrable residue and in solid and semisolid samples.

Total organic acids (TOA) were measured by the chromatographic separation method-504A of Standard Methods (1). However, when a sample contained a high concentration of DNOC or its by-products, the identification of the final titration endpoint was clear due to interference by the yellowish or reddish color of DNOC or its

anaerobic degradation products.

Total organic carbon (TOC) analyses were made in a Beckman Carbonaceous Analyzer Model 915B equipped with total and inorganic carbon channels.

RESULTS AND DISCUSSION

Preliminary shaker reaction results (Table 2) in which sucrose (3.0 g/l) was used as a co-substrate clearly indicated that DNOC was degradable under anaerobic conditions. Over a period of 7 days, the original DNOC concentration of 100 mg/l was reduced to 15 mg/l. Additional results when filtered primary sludge was used as co-substrate provided supporting evidence that DNOC is anaerobically degradable.

DNOC was treated in the anaerobic-aerobic system over influent concentration range of 10 mg/l to 750 mg/l. DNOC was not detected in the anaerobic column effluent with DNOC concentrations as high as 600 mg/l, when the sucrose concentration in the influent was maintained at 3.0 g/l and the hydraulic retention time (HRT) at 3 days.

The effect of the HRT on the performance of the anaerobic column was determined a sucrose concentration of 2.0 g/l, a DNOC concentration of 250 mg/l, and the HRT ranging from 0.17 to 3.0 days. The removal of DNOC by the anaerobic column was $\geq$99.7% even for HRT as short as 0.17 days. Under these operational conditions, no effect of HRT on the removal of DNOC was observed. However, the anaerobic gas production was increased when the HRT decreased at constant influent concentration.

The GC analysis of gas produced in the anaerobic column provided information on the by-products of the anaerobic breakdown process. These analyses were performed when gas production exceeded 3 liters/day. Based on these limited data, the gas composition was 82 to 85% CO_2, 17 to 15% N_2 and less than 1% CH_4. These data indicate that acid-producing bacteria, dominated the microorganisms population in the anaerobic column. The production of nitrogen gas also indicate that a denitrification process occurred in the anaerobic column. Thus the denitrifying microorganisms are responsible for the reduction of the nitro groups that originated from the DNOC molecule, to nitrogen gas.

The continuously-stirred batch reaction results indicated that the degradation of DNOC occurring under anaerobic conditions was a result of the co-metabolism of the nitroaromatic compound and depended upon the presence of a co-substrate. In the presence of the co-substrate sucrose, DNOC was anaerobically co-metabolized. The results presented in Figure 2 demonstrate that the metabolic

TABLE 2. ANAEROBIC SHAKER BOTTLE BATCH TEST RESULTS

Time (days)	Sample*	pH	SS (mg/l)	VSS (mg/l)	TOA (mg/l)	COD_S (mg/l)	TOC (mg/l)	DNOC (mg/l)
0	S	8.43	138	64.0	341	3130	1720	--
	D	8.54	142	69.0	371	3340	1780	100
1	S	6.84	279	197	2150	2550	1430	--
	D	6.89	246	153	1950	2810	1610	78.6
2	S	6.82	245	177	2210	2600	1390	--
	D	6.84	188	123	2230	2780	1530	61.2
3	S	6.70	203	145	2310	2630	1490	--
	D	6.79	172	115	2160	2790	1780	39.8
4	S	6.77	234	156	2700	2560	1360	--
	D	6.85	172	114	2180	2710	1400	21.9
5	S	6.75	130	111	--	2850	1660	--
	D	6.81	89.5	68.5	--	2980	1660	18.0
6	S	6.76	114	93.0	--	3100	1750	--
	D	6.82	143	86.5	2170	3240	1810	15.5
7	S	6.76	88.0	73.5	2600	2790	1320	--
	D	6.83	80.0	56.5	2260	2820	1390	14.6

*S - Sucrose Only (3.0 g/l); D - Sucrose (3.0 g/l) + DNOC (100 mg/l)

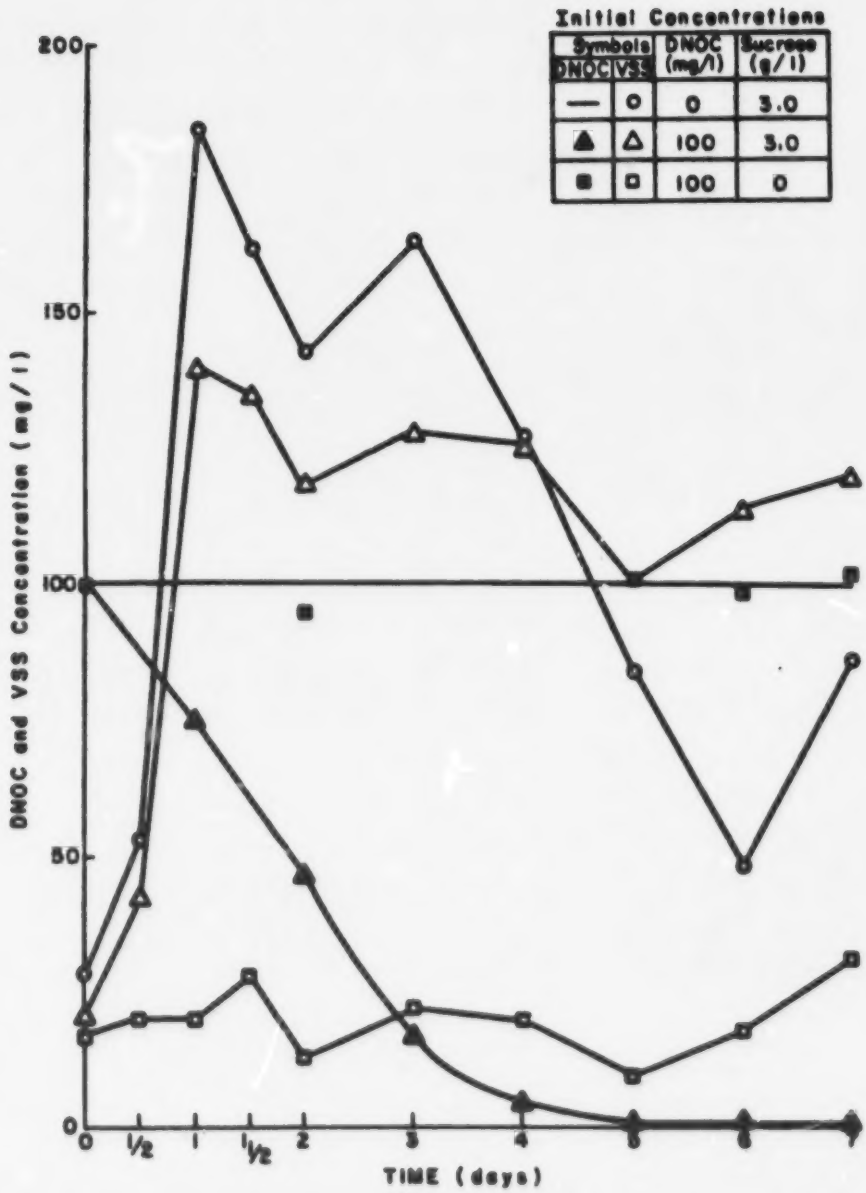

Figure 2. The effect of sucrose concentration on the co-metabolism of DNOC in the continuously-stirred batch test.

TABLE 3. FIRST ORDER DEGRADATION RATE CONSTANTS OF DNOC IN ANAEROBIC CONTINUOUSLY-STIRRED BATCH TESTS

Initial Concs.			$\frac{C_{So}}{C_{Do}}$	First Order Rate Const.	Index of Determin.
Sucrose (g/l)	DNOC (mg/l)	$NaHCO_3$ (g/l)		k_D	R^2
3.0	100	10.7	30	0.0118	0.999
3.0	50	10.7	60	0.0135	0.829
3.0	150	10.7	20	0.00973	0.909
3.0	200	10.7	15	0.00990	0.930
3.0	250	10.0	12	0.00271	0.903
3.0	500	10.0	6	0.00137	0.554
3.0	750	10.0	4	0.000623	0.987
4.5	250	11.25	18	0.00287	0.784
1.5	250	3.75	6	0.000535	0.751
6.0	250	13.75	24	0.00495	0.519
4.5	500	5.0	9	0.00224	0.799
7.5	500	8.75	15	0.00433	0.528
3.0	150	15.0	20	0.00317	0.697
3.0	150	10.0	20	0.00454	0.957
3.0	150	5.0	20	0.00513	0.999
2.0	100	3.5	20	0.000641	0.645
1.0	100	3.5	10	0.000844	0.625
0.5	100	3.5	5	0.000183	0.843

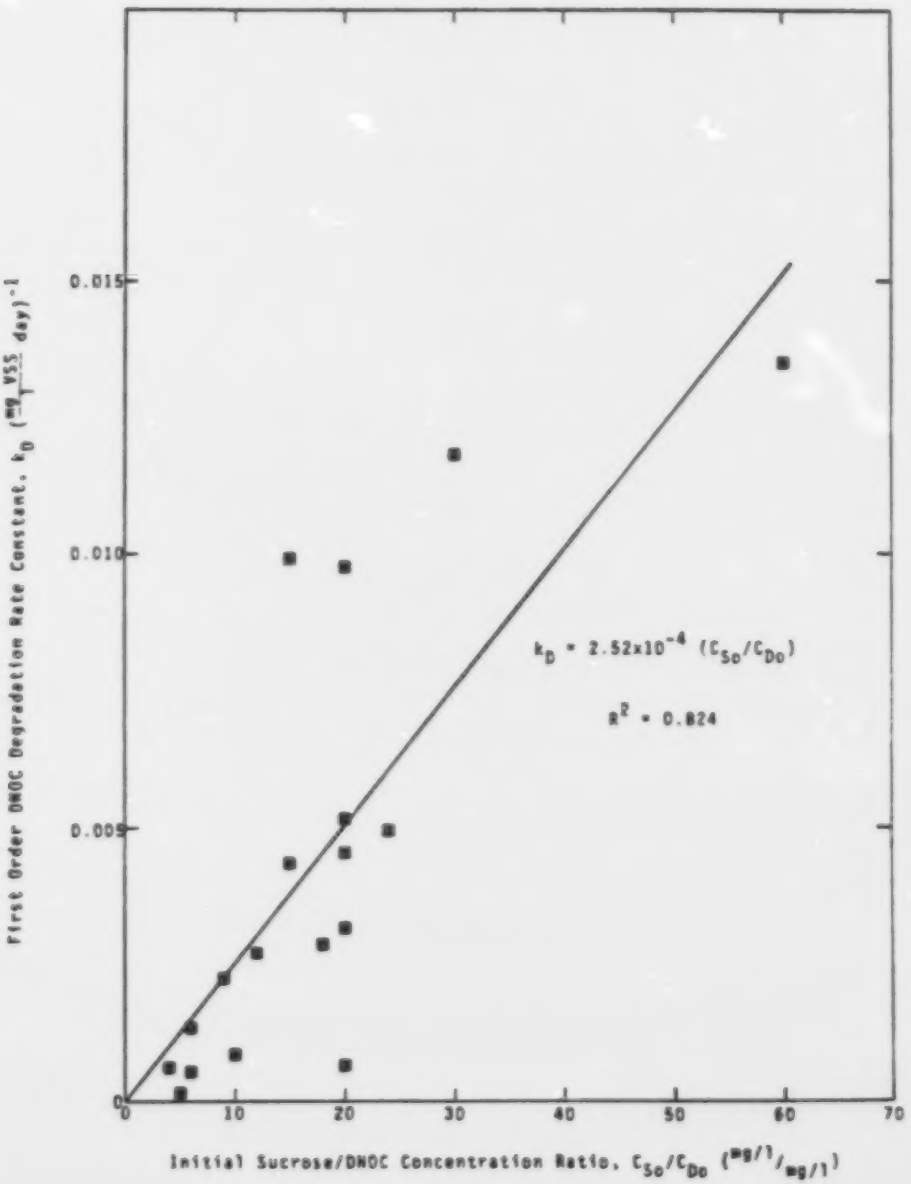

Figure 3. First order rate constant of the degradation of DNOC in continuously-stirred batch tests.

Table 4. FIRST ORDER DEGRADATION RATE CONSTANTS OF SUCROSE IN ANAEROBIC CONTINUOUSLY-STIRRED BATCH TESTS

Initial Concentrations			$\frac{C_{So}}{C_{Do}}$	First Order Rate Const.	Index of Determin.
Sucrose (g/l)	DNOC (mg/l)	$NaHCO_3$ (g/l)		k_D	R^2
3.0	250	10.0	12	0.0105	0.902
3.0	750	10.0	4	0.000294	0.755
4.5	250	11.25	18	0.00391	0.666
1.5	250	3.75	6	0.0101	0.930
6.0	250	13.75	24	0.00237	0.975
4.5	500	5.0	9	0.00335	0.804
6.0	500	6.5	12	0.00693	0.901
7.5	500	8.75	15	0.00209	0.913
3.0	150	15.0	20	0.00588	0.952
3.0	150	10.0	20	0.00880	0.988
3.0	150	5.0	20	0.00961	0.993
2.0	100	3.5	20	0.0442	0.969
1.0	100	3.5	10	0.0189	0.995
0.5	100	3.5	5	0.0277	0.838

process occurs. The microorganisms cannot utilize the DNOC as its sole source of carbon and energy. When there is no sucrose present, the concentration of DNOC remained unchanged. This also resulted in a drastic inhibition of biomass growth; i.e. no change in VSS concentration was observed over time. The DNOC itself is toxic to the anaerobic microorganisms when a readily biodegradable co-substrate is not available.

The anaerobic continuously-stirred batch reactor data were numerically differentiated and fit to the rate equation:

$$-\frac{dC_D}{dt} = k_D C_D X$$

where C_D = concentration of DNOC, in mg DNOC/l,

t = elapsed time since degradation started, in days,

k_D = first order DNOC degradation rate constant in $(\frac{mg\,VSS}{l}\ day)^{-1}$ and

X = biomass concentrations in mg VSS/l.

Table 3 gives the results.

The DNOC degradation rate constant is linear with the ratio of the initial sucrose concentration to the initial DNOC concentration. This relationship is shown in Figure 3.

The degradation of sucrose was modeled using the rate equation:

$$-\frac{dC_S}{dt} = k_S C_S X$$

where C_S = concentration of sucrose, in mg sucrose/l and

k_S = first order sucrose degradation rate constant, in $(\frac{mg\,VSS}{l}\ day)^{-1}$.

Table 4 shows results of this model for the degradation of sucrose in batch tests.

Removal of COD in Anaerobic-Aerobic Pilot Plant

It has been established that DNOC will be degraded or converted under anaerobic conditions in the presenceof sucrose as a cosubstrate. However, this degradation of DNOC, as monitored by GC analysis, was not associated with an appreciable decrease in soluble

COD.

One of the early batch tests resulted in DNOC removal of 85% over a period of 7 days. The removal of DNOC in the anaerobic column was therefore studied under the influent DNOC concentration of 250 and 500 mg/l. The influent sucrose concentration were gradually reduced at fixed hydraulic residency times. These results are shown in Figures 4 and 5.

The performance of the anaerobic column was highly dependent on the influent concentration of the co-substrate, sucrose. A ratio of influent sucrose to DNOC of 2:1 or higher resulted in 95-100% removal (or conversion) of DNOC in the anaerobic process. However, when influent sucrose to DNOC ratio was less than 2:1 the anaerobic microorganisms failed to co-metabolize DNOC.

The degradation of the DNOC molecule in the anaerobic process was not associated with an appreciable decrease in soluble COD. Preliminary data indicated in one case that over a period of 7 days a DNOC removal of 85% was associated with 16% decrease in TOA. In another case a removal of 99.8% of DNOC was associated with only an 11% decreasein soluble COD, while the TOA level increased by 700%. These early data demonstrated that in the presence of sucrose, the anaerobic microorganisms co-metabolized the DNOC molecule and, thereby, caused some initial structural change(s) in or around its benzene nucleus. However, since no appreciable COD reduction occurred in the anaerobic process and a very large increase in TOA occurred, the two main processes which took place in the anaerobic system were: a) the co-substrate, sucrose, was degraded by the anaerobic bacteria to organic acids resulting in a large increase in TOA concentration and an insignificant decrease in soluble COD and b) the structural conversion of the DNOC molecule via cometabolism by the anaerobic bacteria which was dependent on the presence of the co-substrate sucrose.

The reduction in COD levels across the entire anaerobic-aerobic pilot plant was studied under varied conditions of hydraulic retention time in the anaerobic column and varied influent sucrose concentration.

The effect of the hydraulic retention time (HRT) in the anaerobic column on the removal of soluble COD was studied under influent sucrose and DNOC concentrations of 2.0 g/l and 250 mg/l, respectively. The HRT in the anaerobic column was varied from 0.16 days to 3.0 days and the soluble COD removal in the anaerobic column varied from 26% to 3%, respectively. The soluble COD removal achieved in the subsequent activated sludge reactor varied from 24% to 87% resulting in an overall soluble COD removal of 50% to 90%, respectively. The results demonstrated that within the operational conditions of the system, a short HRT in the anaerobic column such as 0.16 days and 0.28 days had an adverse effect on the overall

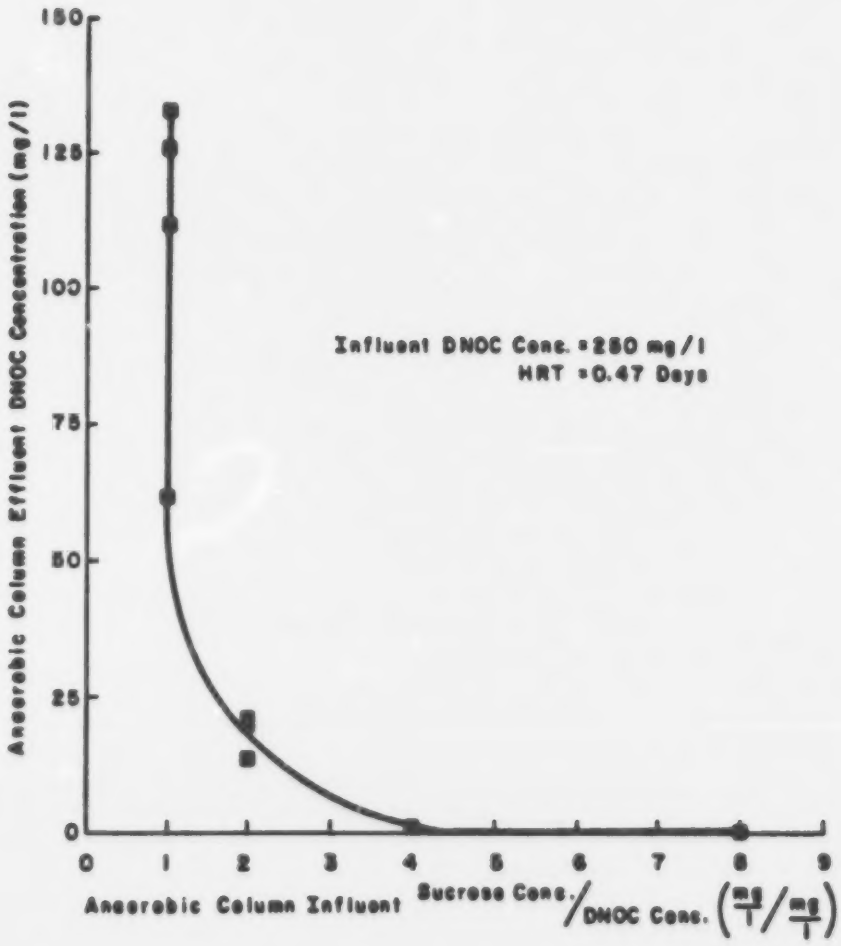

Figure 4. The effect of sucrose/DNOC concentration ratio on the removal of DNOC in the anaerobic column, run 1.

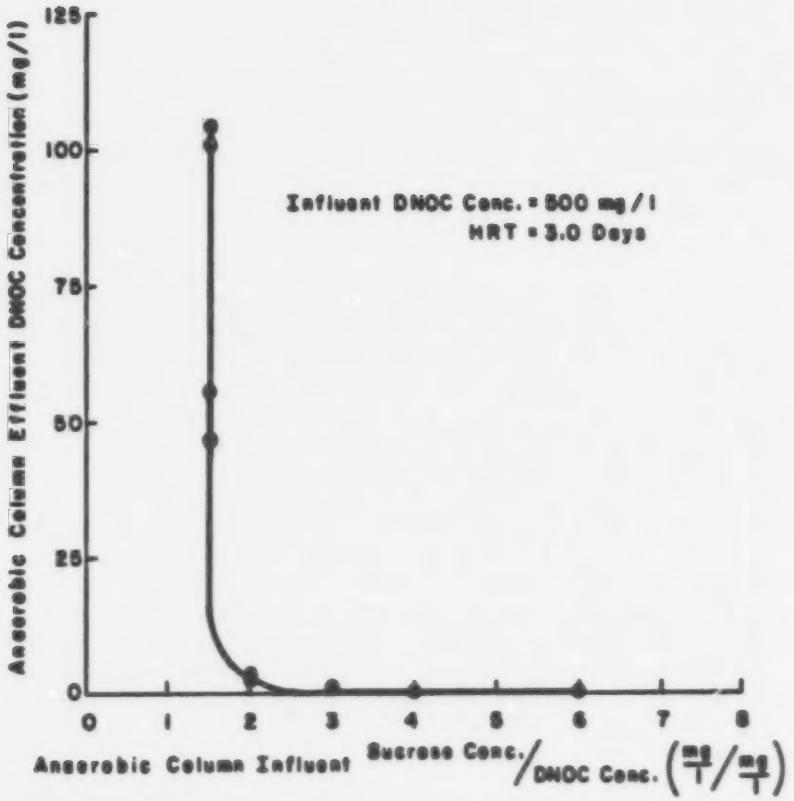

Figure 5. The effect of sucrose/DNOC concentration ratio on the removal of DNOC in the anaerobic column, run 2.

removal of COD, 50% and 70% removal respectively, while a HRT of 0.43 days and longer resulted in overall COD removals of about 90%.

The effect of influent sucrose concentration on the removal of soluble COD through the entire anaerobic-aerobic system was studied under two sets of conditions. The results of these runs are shown in Figures 6 and 7.

When the ratio of influent sucrose concentration to influent DNOC concentration was maintained at 2:1 or higher, the degradation (or conversion) of DNOC in the anaerobic column was between 92% and 100%. However, even when provided with adequate supply of co-substrate, sucrose, to achieve more than 90% conversion of DNOC, the anaerobic microorganisms were unable to decrease the soluble COD by more than 23% in the first set of conditions and 30% in the second set of conditions. The importance of the subsequent aerobic stage in the removal of the remaining high COD is clearly demonstrated. The soluble COD removal that was achieved in the activated sludge process (following an anaerobic DNOC conversion of 90% or higher) ranged from 54 to 71% under the first set of conditions and from 48 to 63% under the second set of conditions.

It is also important to note that the performance of the activated sludge process was adversely affected when the ratio of sucrose:DNOC in the anaerobic column influent was less than 2:1 and reflects the inability of the anaerobic microorganisms to satisfactorily co-metabolize DNOC.

CONCLUSIONS

The effectiveness of the anaerobic-aerobic system in the treatment of DNOC is demonstrated. In contrast to the inability of conventional activated sludge to remove DNOC, the DNOC was completely converted during the anaerobic pretreatment stage. The anaerobic bed effluent DNOC concentration was less than 1 mg/l, with a DNOC concentration in the influent up to 600 mg/l. It was also found that while 100% conversion of DNOC occurred during the anaerobic pretreatment stage, there was less than 25% removal of COD during that stage. In the activated sludge process most of the COD was removed resulting in 80-90% overall COD removal.

The batch kinetic data for the degradation of DNOC and sucrose were modeled by first order kinetics with respect to the DNOC and sucrose concentrations, respectively. The batch rate constant obtained from the DNOC decomposition was linear with respect to the ratio of the initial sucrose to internal DNOC concentrations.

The initial breakdown of DNOC in the anaerobic pretreatment stage was due to co-metabolism. The DNOC itself was toxic to the anaerobic microorganisms if no other, readily biodegradable,

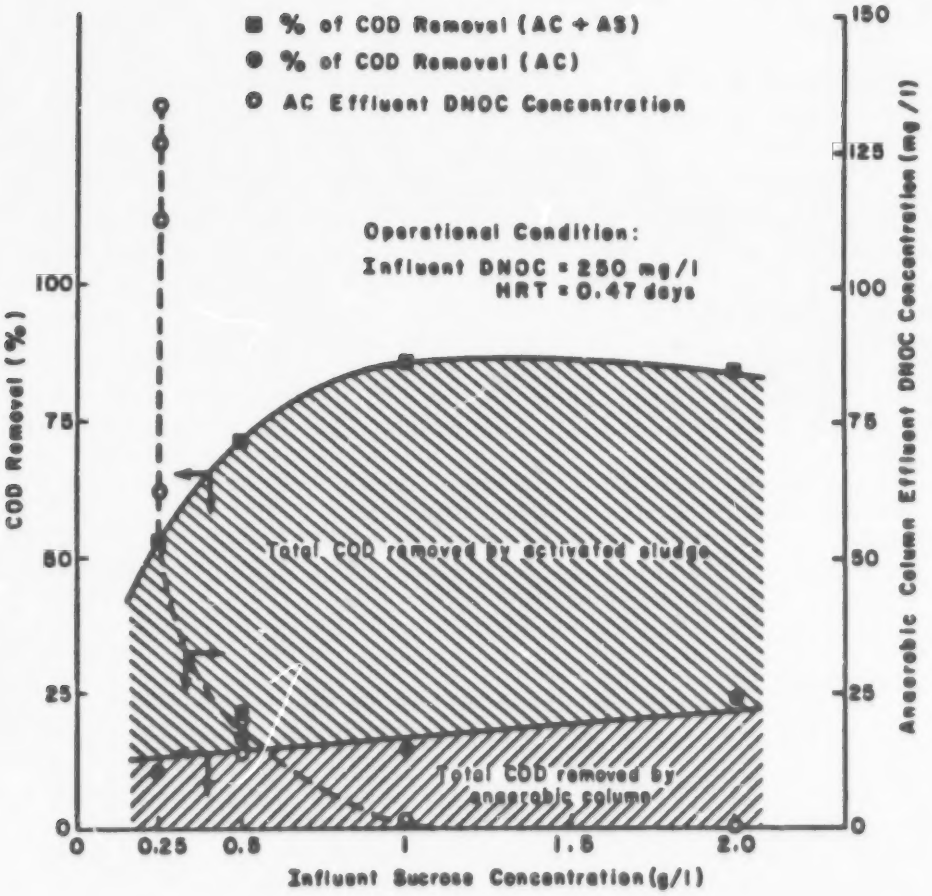

Figure 6. The effect of influent sucrose concentration on the removal of COD by anaerobic column and activated sludge, run 1.

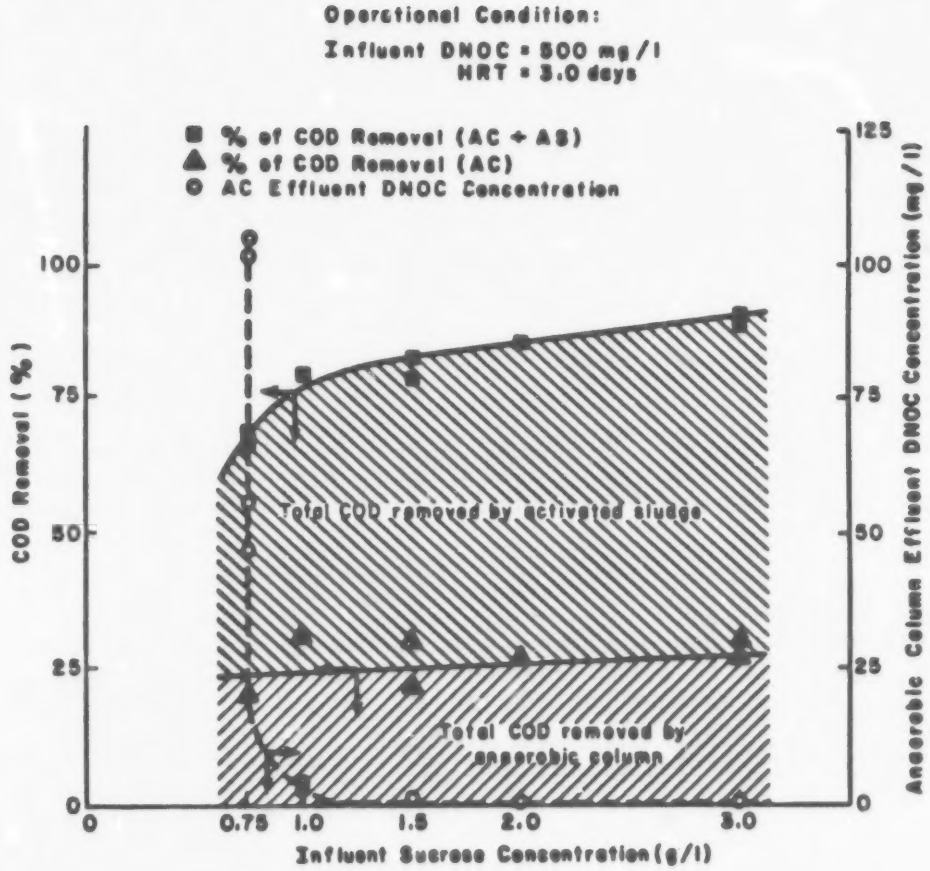

Figure 7. The effect of influent sucrose concentration on the removal of COD by anaerobic column and activated sludge, run 2.

co-substrate was available to them. DNOC cannot be, therefore, used as a sole carbon source by the anaeobic bacteria. Sucrose was used as the co-substrate in this study. The data suggest that there is a relationship between influent sucrose concentration and the performance of the anaerobic pretreatment stage in the conversion of DNOC. The ratio of sucrose to DNOC of 2:1 or higher resulted in a 95-100% conversion of DNOC in the anaerobic pretreamtment stage. The anaerobic microorganisms failed to co-metabolize DNOC when the sucrose to DNOC influent concentration ratio was less than 2:1.

DNOC can be regarded as a model compound representing other degradation-resistant aromatic compounds which, due to their recalcitrancy, present the need for an innovative treatment proces for their removal from wastewater. The high removal rates of DNOC achieved in this treatment process indicated that this treatment process should be evaluated for other recalcitrant phenolic or nitroaromatic priority pollutants.

ACKNOWLEDGEMENTS

This research was supported by grant No. CR810229-01-1 from the U.S. E.P.A. The guidance and assistance of Mr. L.H. Myers and Dr. T. Short are greatly acknowledged.

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PHOSPHORUS REMOVAL FROM AERATED LAGOONS USING ALUM AND FERRIC CHLORIDE - A CASE STUDY

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Phosphorus removal is essential in tertiary treatment of wastewater to prevent lake eutrophication while biological treatment can reduce the total phosphate load on the system to an extent, the process is affected adversely by cold temperatures. Hence this method of phosphorus control cannot be relied upon during winter season. Physico-chemical methods have been tried since a long time in wastewater treatment and they have proved very effective in nutrient control even in cold environment.

Objective of the study

Aerated lagoons remove about 60% of total phosphates by biological action. But there still remains a considerable portion of the phosphate which leaves the treatment plant without any further reduction. An additional problem is caused by dilute wastes which affect the overal treatment of the wastewater. In small sewered communities, aerated lagoons have proved to perform satisfactorily as far as BOD and SS removals are concerned. However, phosphorus and nitrogen removal to the required degree cannot be achieved often, without resorting to some form of physico-chemical treatment. The objective of the present study is to conduct laboratory and on-the-site tests on phosphate removal using essentially two coagulants, namely, Alum and Ferric Chloride. With the results generated over a fairly long period, we should develope simple mathematical models relating the coagulant dosage to nutrient concentration in the wastewater. Secondly, the effect of temperature variations on phosphorus removal is to be examined. Lastly, the choice of phosphorus removal between primary and tertiary levels of treatment is to be evaluated in case of an existing aerated lagoon treatment plant.

Introduction

It is well known that a concentration as low as 1 mg/L of total phosphorus (TP) in the wastewater effluent is sufficient to trigger on the undesirable eutrophication in the receiving stream. Therefore municipal treatment plants have to examine the various effective alternatives for nutrient reduction. Two commonly used methods of phosphorus reduction are:

(a) biological methods and
(b) physico-chemical methods.

The former is slower and sensitive to temperature variations between summer and winter. The latter, namely physico-chemical method of phosphorus removal, is the subject of this on-going study.

Methods and materials

In order to conduct the study an existing aerated lagoon treatment plant is chosen. The wastewater is domestic in character and at present,

removal of phosphorus is through secondary treatment only.

Sampling was conducted in the following points: (1) Influent wet well (2) Polishing pond (3) Effluent weir. The study lasted from January through December. Sampling continued regularly except a few weeks in the month of July due to vacation. Sample collection was daily in the mornings between 9 and 10 AM and alternate days in the evenings. Good samples as well 24 hs composite samples were used in the study. 20 litre polyethylene containers were used to transport the wastewater samples in Coleman Coolers to the laboratory. No preservatives were added as the samples were analysed almost immediately upon arrival in the laboratory. Raw samples were passed through a table top mixer in order to homogenise the constituents. Routine analysis were conducted like the pH, temperature, biochemical oxygen demand BOD, suspended solid (SS), total organic carbon (TOC). Also, various phosphorus modifications like filterable orthophosphate (FOP), filterable phosphate (FP) and total phosphate (TP) were measured.

For tests were conducted in the laboratory and on-the-site also to determine the coagulant dosage for phosphate removal. Two coagulants were chosen namely Alum ($A\ell_2(SO_4)_3$ 14,3 H_2O) and ferric chloride ($FeCl_3$) A typical test non consisted of the following: 6 litres of wastewater samples were taken in 6,1 litre beakers and coagulant was added. After thoroughly stirring for 2 minutes the suspension was allowed to settle. Supernatant liquid was siphoned and analysed. In each case an optimum pH was established after several trials with this optimum pH value, the coagulant dosage was increased gradually till the best floculation and precipitation were obtained.

Standard methods [1] were followed for the determination of various modifications of phosphorus as well as other routine analyses like BOD and SS. In the latter part of the study an auto analyser II was used for phosphate modifications, particularly the orthophosphates.

Results and discussion

We know that a linear relationship exists between phosphorus removal and coagulant dosage as follows [2]

$$\ell n\ (p_r/p_o) = K_p\ M/p_o \qquad (1)$$

where p_r = residual phosphorus concentration, (mg/L)

p_o = initial phosphorus concentration, (mg/L)

K_p = proportionality constant

M = coagulant dosage as metallic ion, (mg/L)

Phosphorus removal at primary level by Alum

Influent samples from the wet well were used to determine the phosphorus removal by Alum precipitation. pH adjustment was not made at first, as we had to compare the values with pH adjustment later. Results obtained over twelve months are plotted in figure 1. Seeing the straight line relationship we also notice at once that temperature variations of wastewater from winter to summer did not influence the precipitation process of filterable phosphorus. The points on the graph can be easily described by a linear equation as follows:

$$\ell n\ [(P_r)_f/(P_o)_f] = -\ 1.43\ A\ell/(P_o)_f + \ell n\ 118.14 \qquad (2)$$

where $(P_r)_f$ = residual concentration of filterable phosphate (FP) (mg/L, P)

$(P_o)_f$ = initial concentration of filterable phosphate (mg/L, P)

$A\ell$ = Alum dosage in (mg/L).

The dotted lines on the figure represent the 5 and 95 percent confidence limits as developed by Viessman [3]. Our results compare very well with those of other authors like Barth et al [4] who have worked on the activated sludge treatment. Table 1 gives a comparison of the results obtained with those of EPA [5].

Table 1: Comparison of ($A\ell$/P) ratios obtained with those recommended by EPA

Percentage reduction of phosphorus	$A\ell$/P ratio	
	Observed value	Recommended value by EPA
75	1.10	1.20
85	1.45	1.50
95	2.20	2.00

The table shows that in case of the aerated lagoons studies, the observed ratios are close to those recommended by regulatory agencies. We also observed that in case of Alum used as a coagulant the filterable phosphate (FP) is more readily precipitated than the filterable orthophosphate (FOP). That is to say, that the polyphosphate which is the difference between the orthophosphate and filterable phosphate reacts preferentially. Our results also indicates that the filterable orthophosphate fraction precipitates only at an ($A\ell$/P) ratio of 0.70 and higher. Below this ratio, the phosphate removal is only that of polyphosphates.

Next, under identical conditions the total phosphorus removal was evaluated. Figure 2 shows the graphical representation of the results. As in the previous case (figure 1) seasonal temperatures did not affect the performance of the coagulant. The following relationship can be formulated in case of the total phosphate (TP) removal:

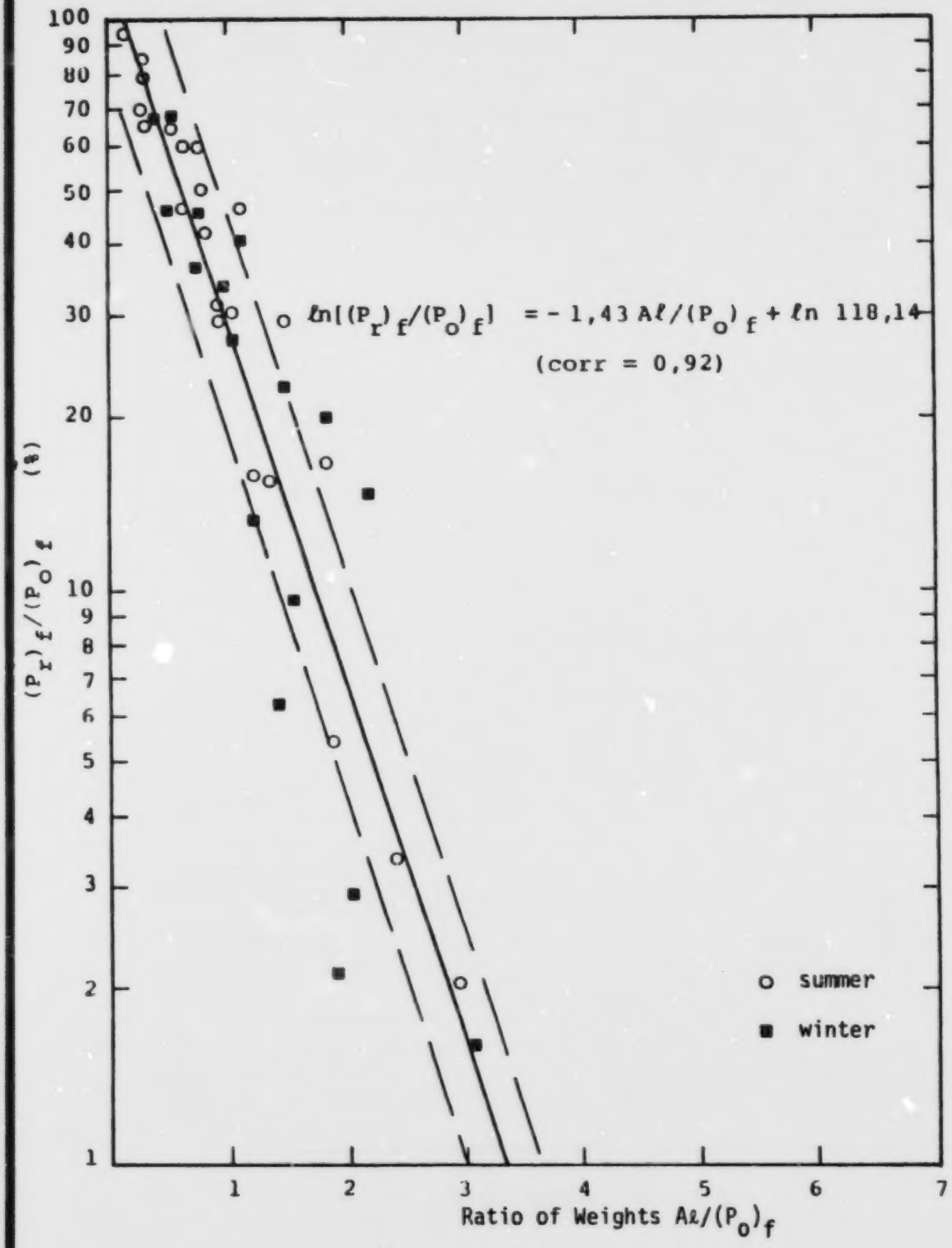

Figure 1. Filterable phosphate removal by Alum at primary level

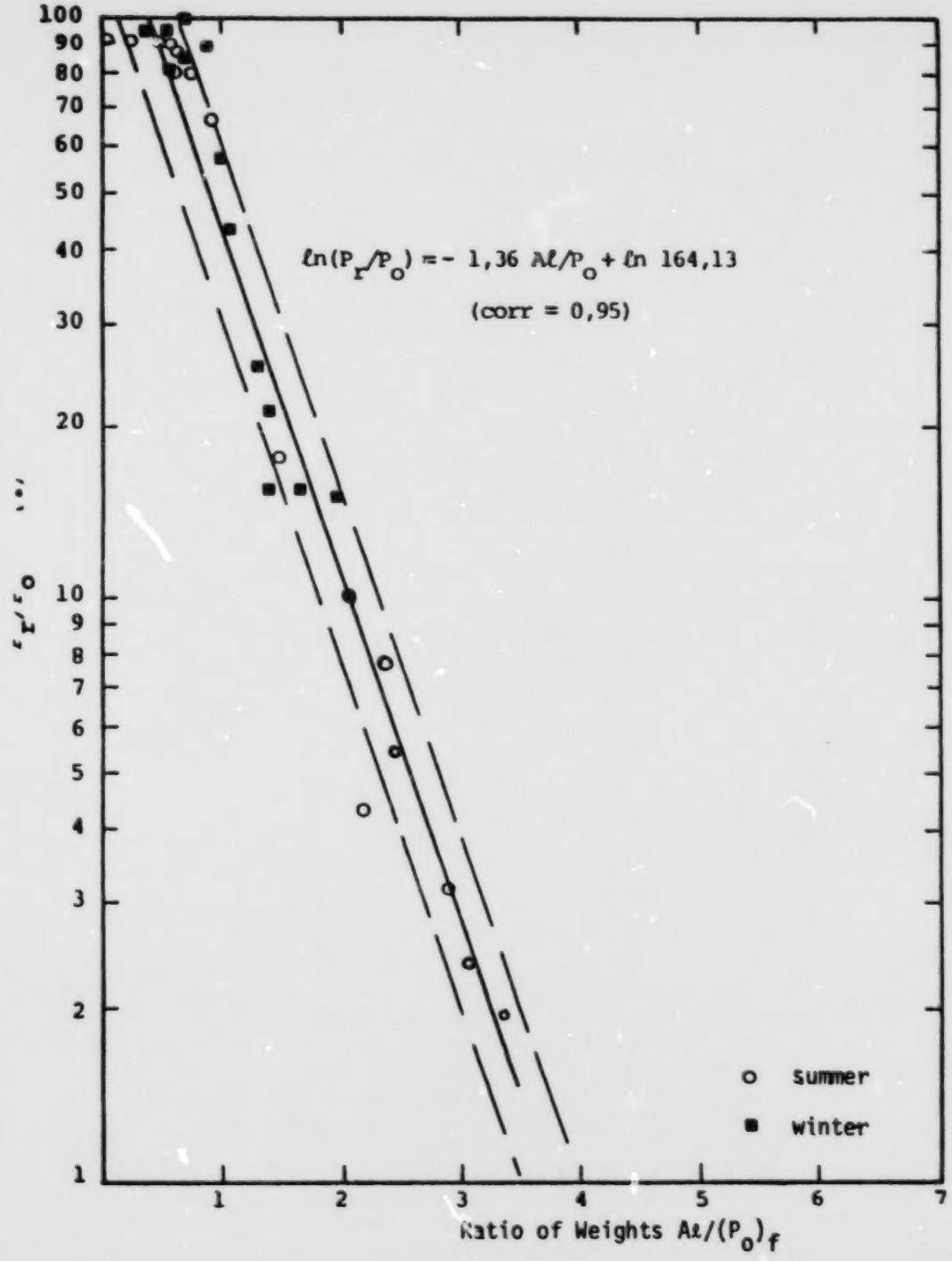

Figure 2. Total phosphate removal by Alum at primary level

$$\ell n\ (P_r/P_o) = -\ 1.36\ A\ell/P_o + \ell n\ 164.13 \qquad (3)$$

where P_r = concentration of residual total phosphorus (TP) (mg/L,P)

P_o = concentration of initial total phosphorus (TP) (mg/L, P)

$A\ell$ = metallic cation dosage, (mg/L).

Here is a comparative table (table 2) showing the performance of Alum for removal of filterable and total phosphate.

Table 2: Comparison of efficiency in removal of filterable and total phosphate by Alum

% Efficiency	Aℓ/P ratio Filterable phosphate	Total phosphate
90	1.70	2.05
95	2.20	2.55
99	3.35	3.75

From the above results we can note that Alum reacts preferentially with filterable phosphate. A ratio of Aℓ/P (by weight) equal to 0.10 is enough to start the precipitation in case of filterable phosphate. For total phosphate removal a minimum of Aℓ/P ratio equal to 0.50 is the precondition. It was observed that a small dosage of coagulant increases the non-filterable phosphate concentration in the bulk solution. Increasing the coagulant dosage at this point, one notices an increase in the phosphate concentration in suspension. This concentration reaches a level where the precipitate formed earlier is resuspended and continuing the dosage leads to a agglomeration of the precipitate gradually. At the end , the concentration of total phosphate in the supernatant decreases as a result of precipitation. The practical use of figure 2 is that it enables the determination of coagulant dosage in terms of percentage removal of phosphorus. For example, for 90% removal of total phosphorus (TP) it is necessary to dose the Alum more than twice the concentration of total phosphorus in the wastewater. In this particular treatment plant we have an annual average phosphorus concentration of 4.80 mg/L which means that we need 4.80 x 2.00 = 9.60 mg/L aluminum = 123.44 mg/L of Alum. To obtain 1 mg/L total phosphate in the effluent, a dosage of (3.80 x 2 x 11.11) = 84.44 mg/L of Alum.

The above experiments were repeated after adjusting the pH to between 6.50 and 5.75. We see in figures 3 and 4 that there is less of dispersion of the points about the straight line and that the confidence limits are narrower. The stability of the results is also reflected by the high degree of correlation in each case. Here are the two equations describing the straight line relationship:

At pH 6.5:

$$\ell n\ [(P_r)_f/(P_o)_f] = -\ 1.57\ A\ /(P_o)_f + \ell n\ 157.52 \qquad (4)$$

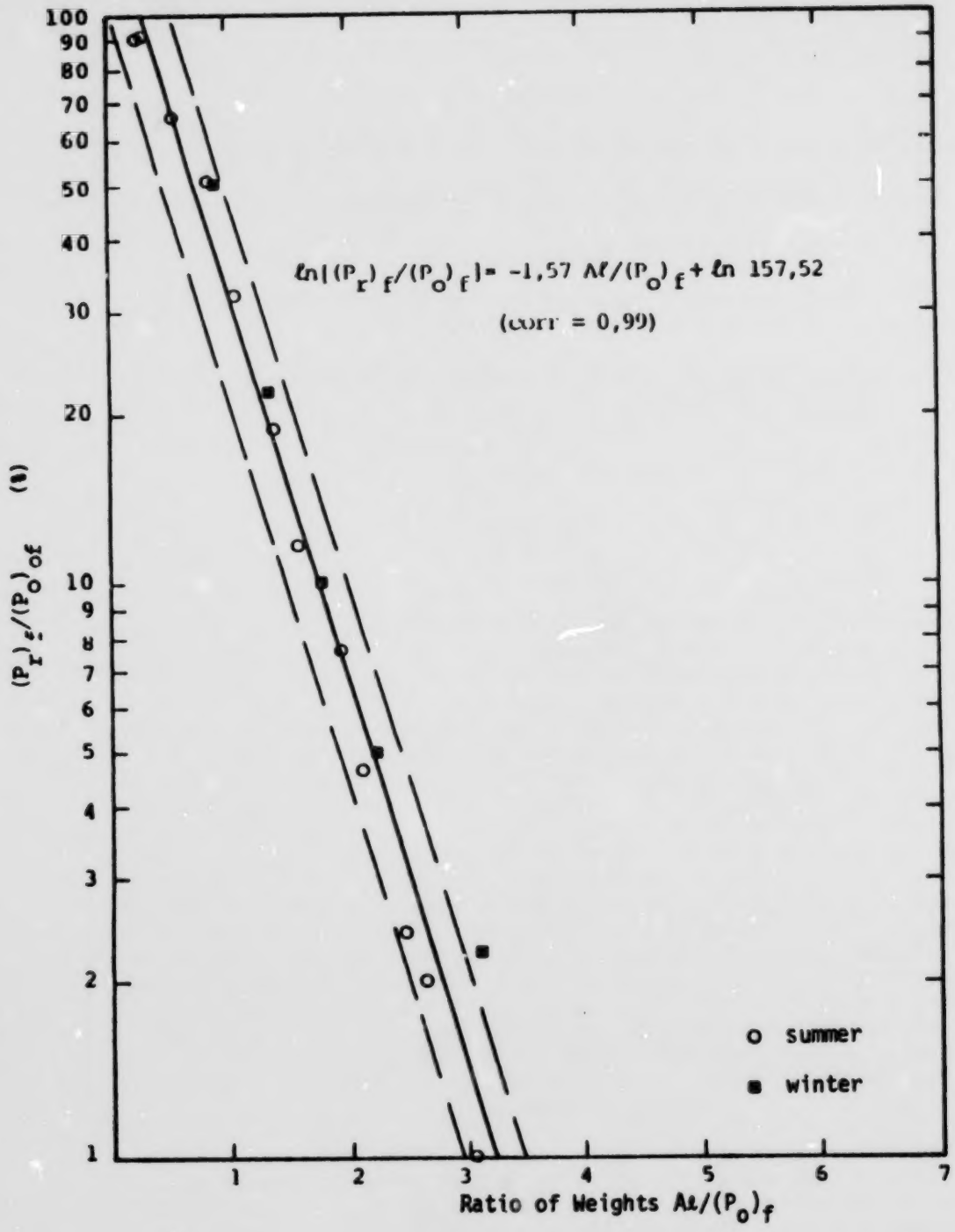

Figure 3. Filterable phosphate removal by Alum at ph 6.50, primary level

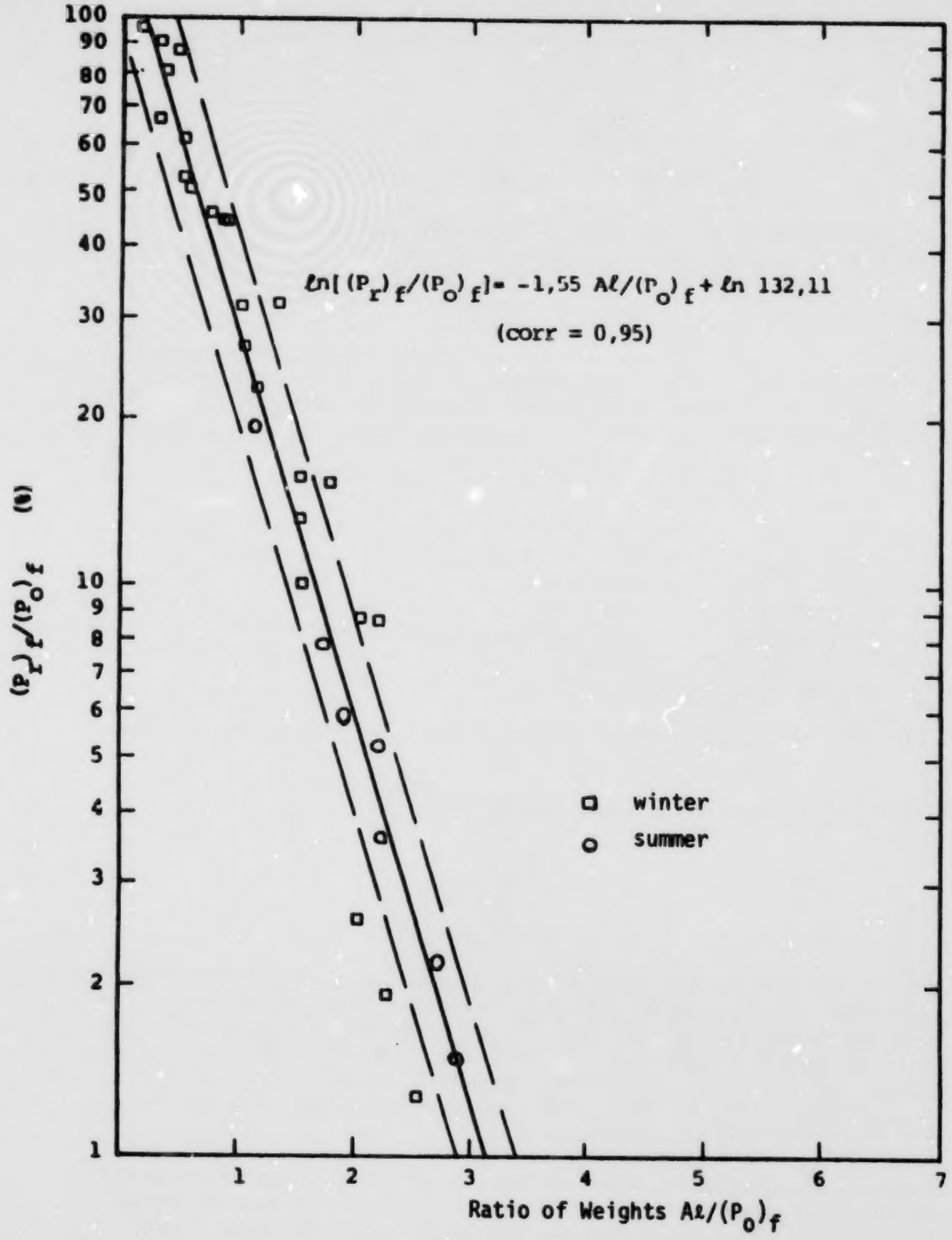

Figure 4. Filterable phosphate removal by Alum at ph 5.75, primary level

At pH 5.75:

$$\ell n\ [(P_r)/(P_o)] = -\ 1.55\ A\ell/(P_o) + \ell n\ 132.11 \qquad (5)$$

From the above results we also noticed that for 99% removal of filterable phosphorus, a reduction of pH up to 6.50 results in a decrease in the Aℓ/P ratio from 3.35 to 3.23. Further lowing of pH to 5.75 the ratio drops from 3.23 to 3.20 only. Economically, however, it would not be worthwhile to lower the pH to 5.75 because of the corrosion and related problems of low pH. Encouraged by the above results, we proceeded to evaluate the total phosphorus removal at the corrected pH value. Similar tendency to flocculate at pH of 6.50 enabled the formulation of the following equation corresponding to figure 5.

$$\ell n\ [(P_r/P_o)] = -\ 3.27\ A\ell/P_o + \ell n\ 654.09 \qquad (6)$$

From the figure, we can see that above an Aℓ/P ratio of 1.70 the coagulant addition did not help any precipitation. On the contrary the precipitate flocs wave resuspended resultives in a higher residual phosphorus concentration in the bulk solution which is not a desirable solution. Isgard [6] obtained the same result at an Alum dosage of 97 mg/L. Here is a table showing the influence of pH on the total phosphate reduction at primary level.

Table 3: Influence of pH on reduction of total phosphorus by Alum

Percentage removal	Ratio Aℓ/P	
	Raw wastewater (pH 7.80)	pH 6.5
90	2.05	1.30
95	2.55	1.50
99	3.75	1.85

From the table, we may observe that lowering pH improve the coagulation only to a certain degree for a 90% reduction of total phosphate, an Aℓ/P ratio of 1.30 at a pH of 6.50 is sufficient, compared with that of 2.05 without pH adjustment. That is, a reduction of ≃ 35% of Alum dosage for an initial phosphorus concentration of about 4.40 mg/L. If we increase the ratio above 1.70 then the overdosage problem is initiated and the precipitation hampered.

A series of experiments were run at the site of the plant as well as in the laboratory in order to determine whether phosphorus removal at secondary level of treatment was any better than at primary level. Samples were collected from the polishing pond and effluent weir were tested during a one year period.

Phosphorus removal by Alum at secondary treatment level

Knowing from the foregoing experience that a pH adjustment was very beneficial, we proceeded to study the Alum coagulation at adjusted

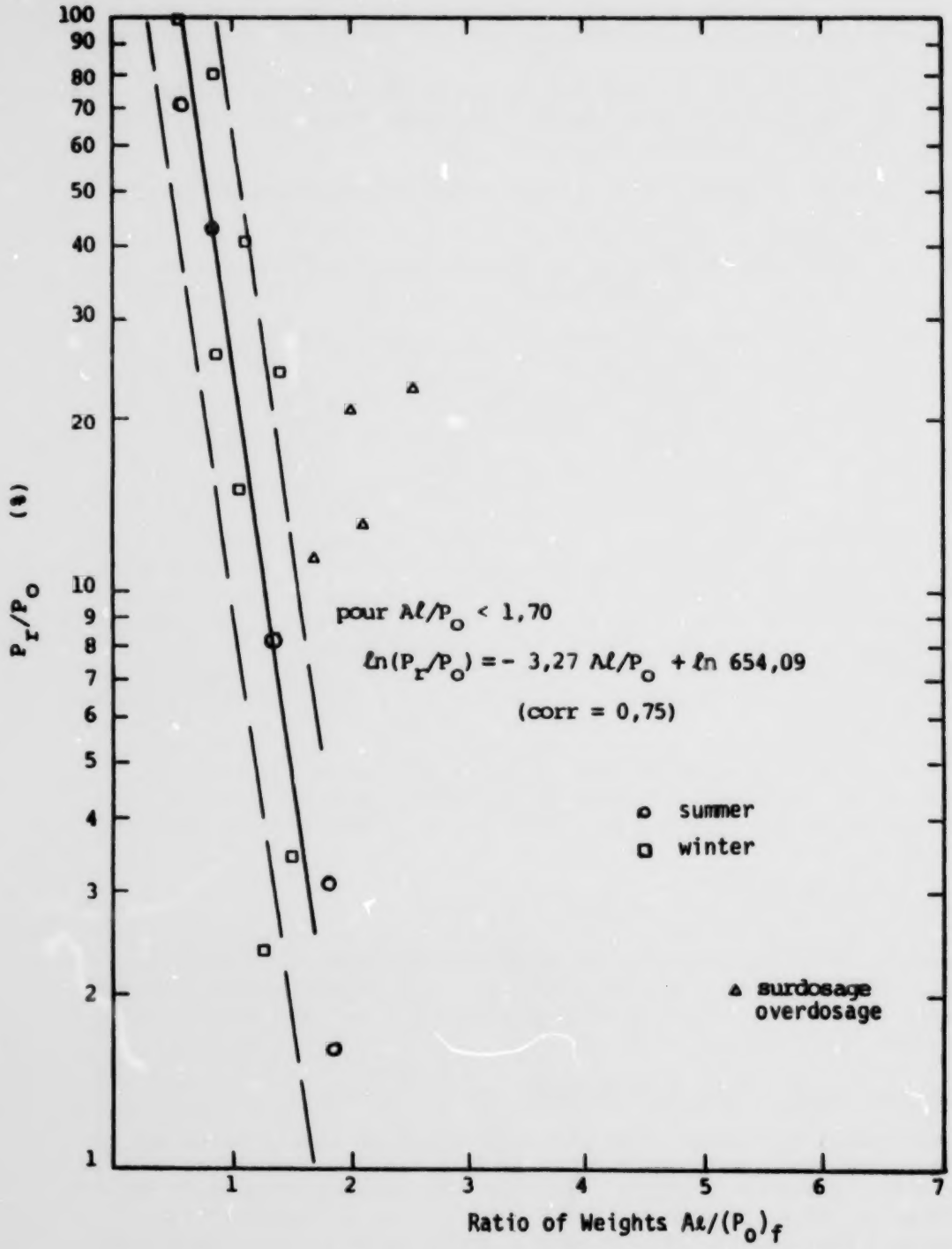

Figure 5. Total phosphate removal by Alum at ph 6.5, primary level

pH value. However, those were two values of pH namely 6.5 and 5.5 where the coagulation seemed identical although the better choice was not difficult to recognise, namely at 6.50.

Table 4 gives a comparison of two pH values on the phosphate precipitation.

Table 4: Influence of pH value on the phosphate precipitation at secondary level of treatment

Efficiency	Effluent treated without pH adjustment	Ratio of Aℓ/P	
		pH 6.50	pH 5.50
90	2.15	1.80	1.69
95	2.80	2.30	2.11
99	7.60	3.50	3.15

From the table we note that by lowering the effluent pH (7.50) to 6.50 the Aℓ/P ratio reduce from 7.60 to 3.50 for 99% removal of total phosphorus. Addition lowering from 6.50 to 5.50 results in a Aℓ/P ratio reduction by 0.35 only. The margin in is even narrower as the percentage efficiency is lower. Hence for all practical purposes it may not be worthwhile attempting to lower the pH from 6.50 to 5.50. Comparing the results obtained of phosphorus (TP) removal at primary level of treatment to that at secondary level we observed that the efficiency at secondary level is lower than at primary. For example, at primary level an $A\ell/P_o$ ratio of 1.50 was enough to achieve almost 97% removal of total phosphorus. For the same efficiency to be achieved during secondary level the Aℓ/P ratio had to go up to 4.50. That is almost three times the dosage of Alum.

Phosphorus removal by Ferric Chloride ($Fec\ell_3$)

Ferric chloride is used often in wastewater treatment for effective coagulation at primary and secondary level. A choice between Alum and ferric chloride for treating aerated lagoon effluent was examined at the same levels of treatment.

At primary level after pH adjustment

As was learnt in case of Alum, establishing an optimum pH value was our primary concern. This was found to be 4.75 for both primary as well as secondary degree of treatment. Authors like Bernard [7] have found optimum pH values between 4.0 and 5.0 in case of activated sludge process. Needless to say that such a low pH value is very impractical on a prototype installation. Figures 6 and 7 show the filterable phosphate removals at primary level at a pH of 6.0 and 4.75. There is not an appreciable difference between the two. The corresponding equations are given by:

$$\ell n\ [(P_r)_f/(P_o)_f] = -\ 60\ Fe/(P_o)_f + \ell n\ 156.39 \qquad (7)$$

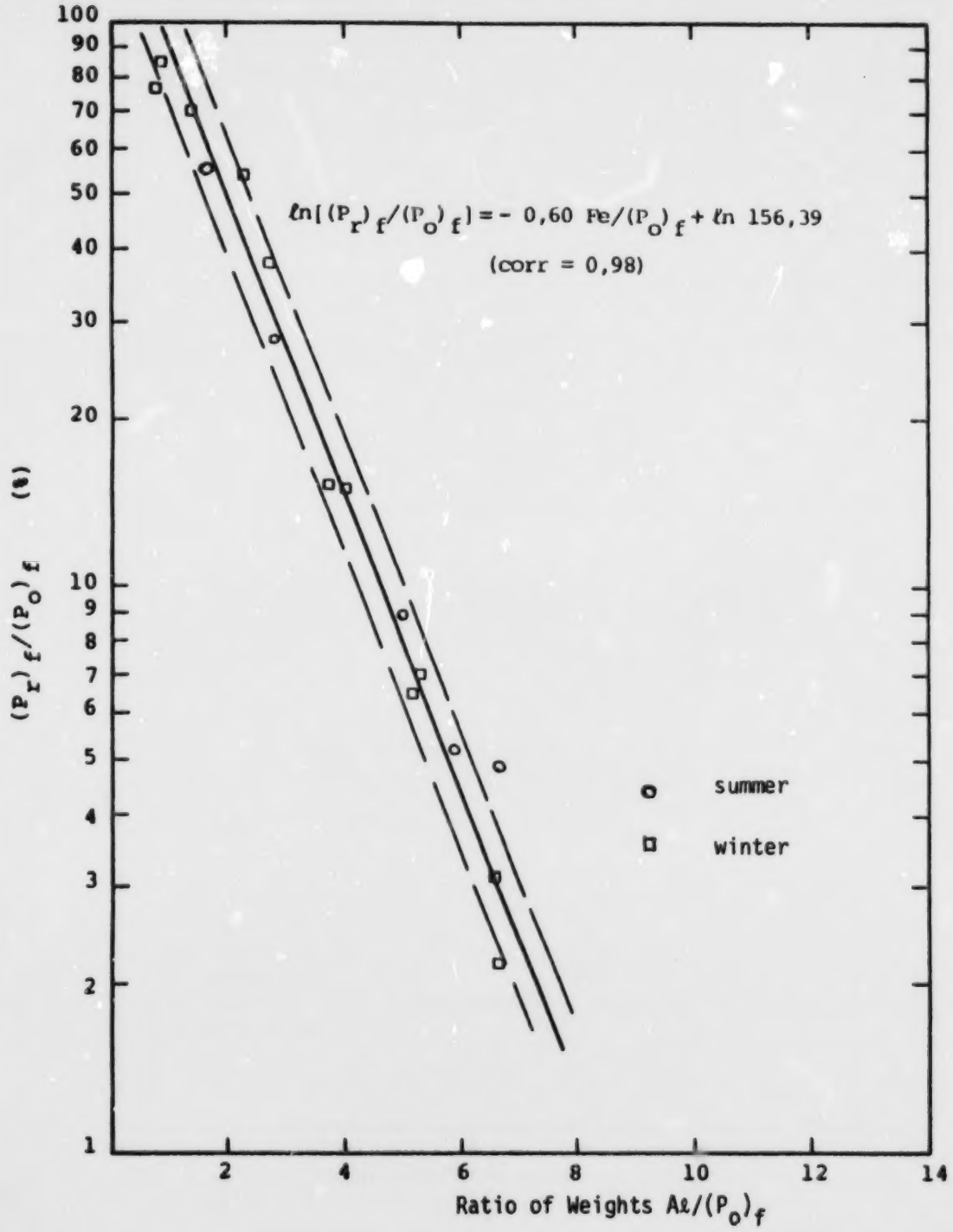

Figure 6. Filterable phosphate removal by ferric chloride at pH 6.0, primary level

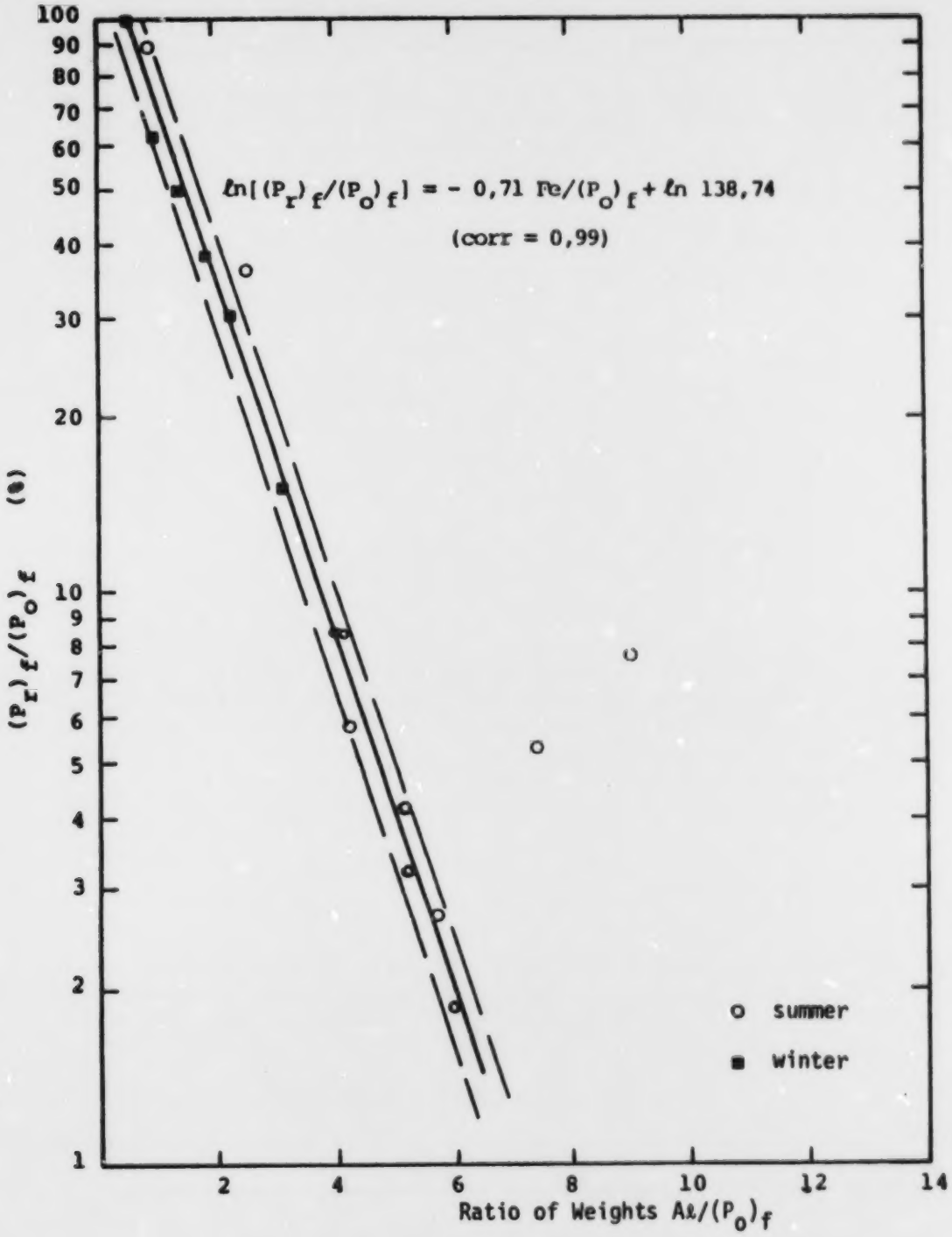

Figure 7. Filterable phosphate removal by ferric chloride, at pH 4.75, primary level

at a pH of 6.0 and

$$\ell n\ [(P_r)_f/(P_o)_f] = -\ 0.71\ Fe/(P_o)_f + \ell n\ 138.74 \qquad (8)$$

at a pH of 4.75.

At secondary level of treatment

Secondary effluent was treated with ferric chloride exactly the same way as in the previous case. The two pH values tried were 6.0 and 4.50 respectively. Figure 8 shows the straight line relationship obtained in case of total phosphorus removal at pH of 6.0.

The equation describing the above relationship is

$$\ell n\ [(P_r)/(P_o)] = -\ 0.26\ (Fe/P_o) + \ell n\ 135.27 \qquad (9)$$

Here is a table of comparison between Alum and ferric chloride as coagulants.

Table 5 : Comparison of Alum and ferric chloride as coagulants in the removal of total phosphates

Removal efficiency	Ratio by weight			
	Alum		Ferric chloride	
	primary	secondary	primary	secondary
90	2.05 (2.36)*	2.80(3.22)	5.30(2.94)	10.20(5.67)
95	2.55 (2.93)	6.25(7.18)	6.80(3.78)	12.80(7.17)
99	3.75 (4.31)	-	10.20(5.67)	-

* Molar ratio

From the above table we can see that as a coagulant for total phosphate removal ferric chloride is less efficient than Alum. The molar ratios being higher in case of ferric chloride than Alum mean greater dosage of the chemical and consequent higher costs.

Summarising the experiments described above the following conclusion can be drawns:

(1) Simple equations relating coagulant dosage to P removal were developed for practical application in case of existing aerated lagoons contemplating nutrient removal.

(2) Alum reacts preferentially with poly-phosphate and not with ortho-phosphate in the precipitation process.

(3) Seasonal temperature variations of air and wastewater have no significant effect on the coagulation and flocculation process.

(4) The molar capacity of Alum for precipitating orthophosphate fraction is greater at tertiary stage compared to primary.

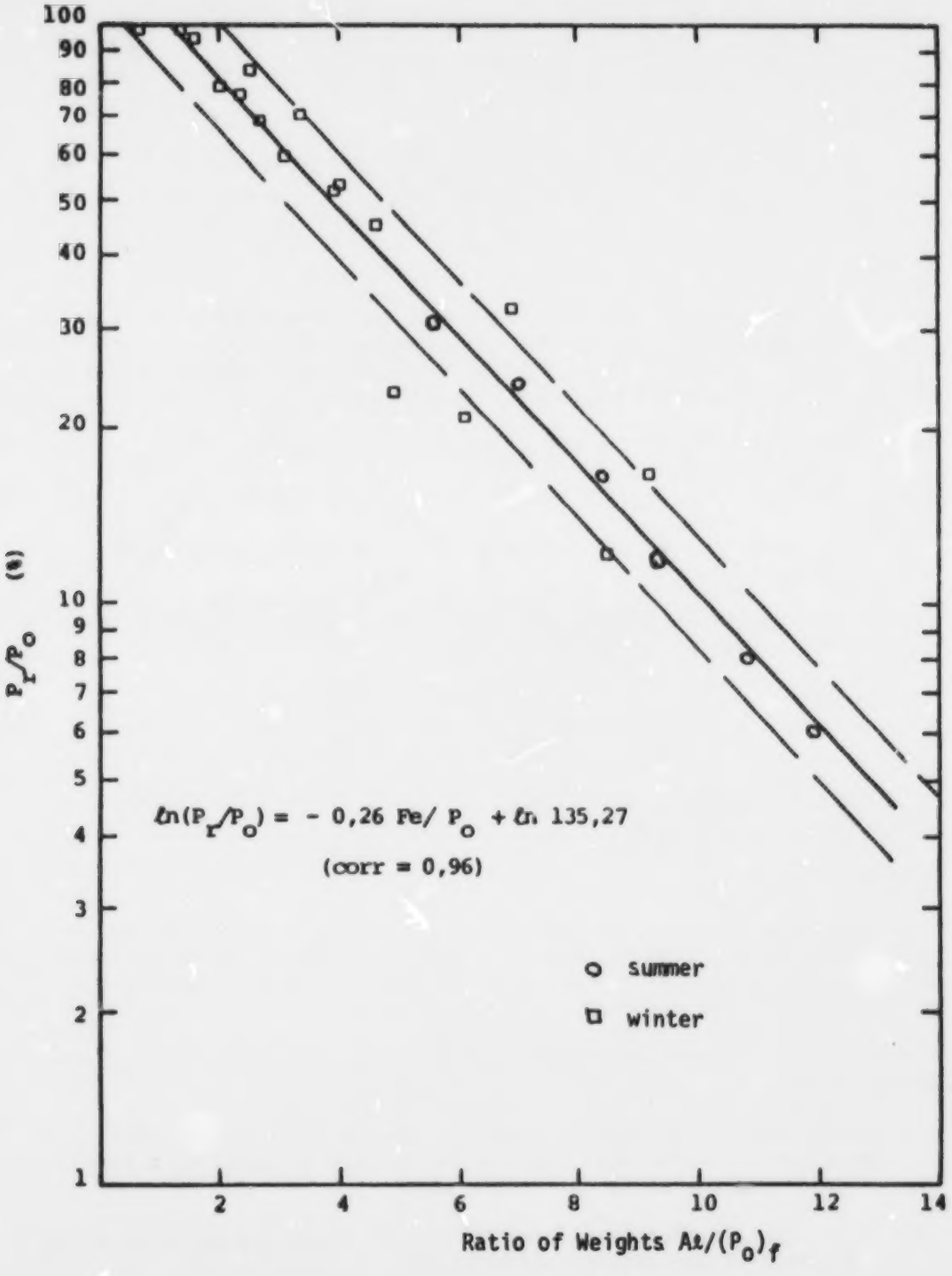

Figure 8. Total phosphate removal by ferric chloride at pH 7.50, secondary level

(5) At the tertiary level, the phosphorus removal involves a higher coagulant dosage than for the same degree of removal at the primary stage of wastewater treatment.

(6) For low concentrations of filterable phosphates, ferric chloride is a better coagulant than Alum as its molar capacity is higher than that of Alum, the only prerequisite being an optimum pH.

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[6] Isgard, E., Chemical Methods in Chemical Method in Present Swedish Sewage Purification Techniques, Allied Chemical, Ltd., Publication No 366-500-1-72 (1972).

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REVERSE OSMOSIS TREATMENT OF COAL-FIRED POWER PLANT WASTEWATER

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ABSTRACT

A study was carried out to investigate the feasibility of applying 'seeded' reverse osmosis (RO) to treat fly ash leachate. RO flux rates for water averaged about 0.89 $m^3/m^2 \cdot d$ operating at 5.5 MPa applied pressure, with overall solute rejection efficiencies of about 97% (TDS) and a volume reduction factor of ~20:1. Product water (permeate) quality was found to be excellent. RO membrane fouling was effectively avoided: gypsum precipitated from saturated solution as finely suspended nuclei that grew to a size removable by in-line centrifugation. Depending on scale and mode of operation, total treatment costs were estimated to range from about $3 to $25 per m^3 of product water.

1. INTRODUCTION

Ontario Hydro has a legal contract for disposal of coal fly ash at Domtar Quarry, Mississauga. The fly ash results from combustion of pulverized U.S. bituminous coal at nearby Lakeview Generating Station. One condition of the contract is that Ontario Hydro provide a contingency plan for treatment of fly ash leachate should it ever be found to be necessary. Since placement of fly ash commenced in December, 1981, there has been no such need. Nevertheless, an exploratory technical and economic feasibility study of treatment of fly ash leachate was undertaken by Ontario Hydro Research Division, considering the further possibility for treatment of similar wastewaters.

Preliminary cost estimates were made of evaporation, a broad spectrum treatment considered to be a fairly certain means of removal of any impurities that might be present. Since evaporation was found to be expensive, it was considered worthwhile to examine the feasibility of applying reverse osmosis (RO), also a broad spectrum water treatment method, which is generally less expensive than evaporation.

It was anticipated that during RO treatment of fly ash leachate the solubility product for calcium sulphate would at some point be exceeded and that gypsum precipitating from the saturated (or supersaturated) solution could, if unchecked, rapidly blind the membrane surfaces, with severe loss of transmembrane flux rate for purified water, and reduced rejection of dissolved impurities. It was therefore proposed to study 'seeded' RO, using tubular cellulose acetate membrane modules (internal flow configuration) whereby gypsum crystallizing out would preferentially do so on growing nuclei forming in suspension rather than on the membrane walls. (More accurately, this might be termed 'self-induced seeded RO', since, during volume reduction, nucleation takes place spontaneously *in situ* - without preliminary introduction of seed crystals of extraneous origin such as has been proposed elsewhere/1/. For brevity, the term 'seeded' RO is used here).

2. EXPERIMENTAL, RESULTS AND DISCUSSION

2.1 'SEEDED' RO ('FEED-AND-BLEED' MODE)

2.1.1 Equipment

An experimental RO test loop was set up as shown schematically in Figure 1, to determine the technical feasibility of converting fly ash leachate by RO into:

(i) a purified stream (permeate), which could be discharged (or perhaps utilized) on-site

and (ii) a low-volume concentrate which could be either (a) used on-site for fly ash dust suppression or (b) economically transported off-site for disposal.

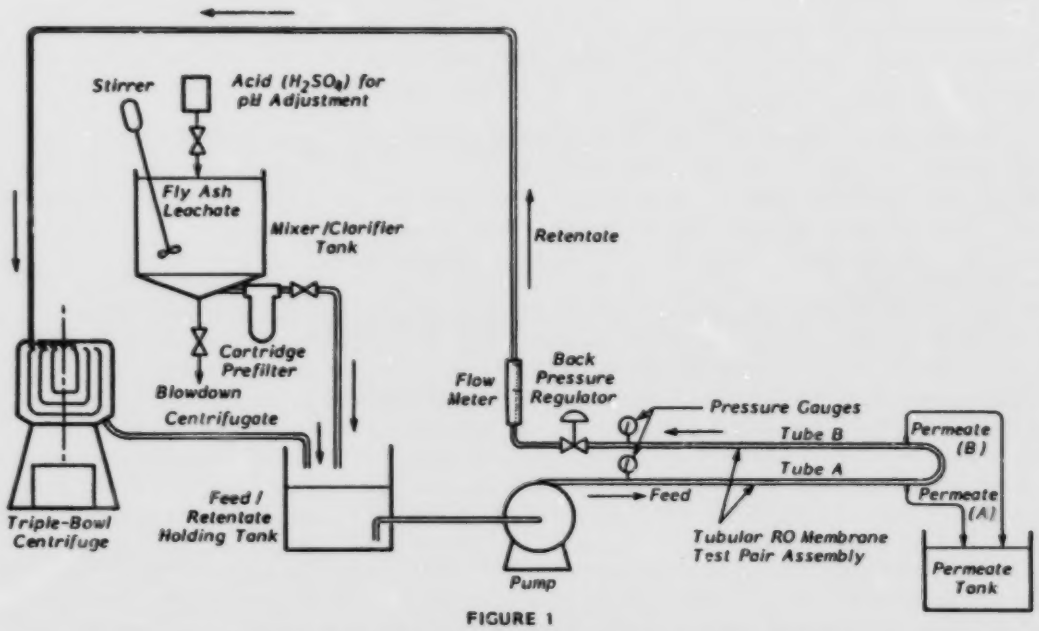

FIGURE 1

SCHEMATIC OF RO BATCH TREATMENT FACILITY FOR TREATMENT OF FLY ASH LEACHATE (NOT TO SCALE)

The RO pilot unit incorporated a tubular cellulose acetate membrane test pair assembly supplied by Abcor, Inc. The cellulose acetate membranes, each 3.05 m long and 1.27 cm ID, were supported inside stainless steel tubes, joined in series by a single U-shaped turn-about end-fitting, representing the simplest repeat unit of the standard Abcor RO production module, which contains 30 tubular membranes. The effective area of an individual membrane tube is about 0.122 m^2. The back-pressure regulator chosen was a Whitey straight-through plug valve of stainless steel construction equipped with a

replaceable teflon valve seat. (To avoid plugging of the back-pressure regulator by gypsum crystals expected to form in the fly ash leachate during RO operation, needle valve designs were avoided.)

An in-line centrifuge was used to remove oversize gypsum particles from suspension in the RO retentate during volume reduction, and prevent their accumulation in the RO system. The centrifuge was a triple-bowl Donaldson unit rated at 0.75 kW, with a nominal fixed rotational speed of 700 rpm.

Fly ash leachates derived from bituminous coal sources are known/2/ to contain relatively high levels of calcium and sulphate (both divalent ions) compared with sodium, potassium and chloride (monovalent ions). Since di- and polyvalent ions generally have higher RO rejection efficiencies/3/ compared with monovalent ions, a lower-rejection membrane was selected for the RO test than would be necessary for effective retention of monovalent ions if they alone were present in solution. There is also the attendant advantage of higher flux rates intrinsically associated with lower-rejection membranes at a given applied pressure.

The membrane chosen for the 'seeded' RO test with fly ash leachate was the 'Abcor' Type 190 tubular cellulose acetate membrane, which has a nominal flux rate for water of 0.81 $m^3/m^2 \cdot d$ and a NaCl rejection efficiency of 90% operating at 4.1 MPa applied pressure.

Temperature control in the feed holding tank was maintained during the test at about 24.5 C by a 'Blue M' stainless steel coil temperature regulator, backed up by a simple water-cooled coil for cooling at low residual volumes when the regulator coil was incompletely immersed.

2.1.2 Pretreatment

A batch of fly ash leachate was made up for use in the RO test by vigorously stirring fresh Lakeview Generating Station fly ash with deionized water for ~16 h at a liquid:solid weight ratio of 4:1. After settling overnight, the supernatant was siphoned off and filtered through a 20 µm cartridge filter. The solution at this stage had specific conductivity 304 mS/m and pH 9.5. Since the chemical stability of cellulose acetate RO membranes is generally accepted as optimal/2/ in the operating range pH 3-8, the leachate was adjusted to pH~7 by adding H_2SO_4. A small amount (about 7g) of a gelatinous greyish-white precipitate formed. After settling and removing the solids, the neutralized leachate was given a polishing filtration through a 20 µm cartridge filter and transferred to an RO feed holding tank constructed of stainless steel. The neutralized RO feed had a volume of 338 L, with a specific conductivity of 260 mS/m, at pH 7.3. Samples of the fly ash leachate, taken before and after pH adjustment, were submitted for detailed chemical analysis, together with a sample of the dried precipitate. Chemical analyses are given in Table I. The greyish-white precipitate appears to consist mainly of calcium sulphate and aluminum hydroxide with minor amounts of the hydroxides of gallium, iron, copper and zinc. There is evidence also of the presence of unburnt carbon fines as reflected by the high level of CO_2 (measured as 'carbonate') formed during analysis of the solid substance by 'Leco' combustion.

The only appreciable change in the leachate composition after pretreatment (Table I) is an increase in sulphate, accompanied by reduced alkalinity and a decrease in aluminum. The amount of most trace metals removed in the precipitate is insignificant.

TABLE I

ANALYTICAL DATA FOR PRETREATMENT OF FRESH LAKEVIEW FLY ASH LEACHATE

Quantity	337 L	0.007 kg	338 L
Stream	RO Feed Before Neutralization (mg/L)	Precipitate After Neutralization (mg/kg)	RO Feed After Neutralization (mg/L)
pH	9.5		7.3
Sp. Cond (mS/m)	304		260
Alkalinity OH^-	40		<5
as $CO_3^=$	92	290,000	<5
$CaCO_3$ HCO_3^-	<5	-	27
SO_4	1620	168,000	2700
Cl	13	80	9.3
NO_3	-	-	-
Ca	580	60,000	580
Mg	0.32	-	0.33
Na	91	1,700	85
K	62	2,100	59
Ag	<0.0002	-	<0.0002
Al	8.6	220,000	<0.5
As	0.01	900	0.02
B	23	-	22
Be	<0.0002	-	<0.0002
Ba	<1	370	<1
Br	0.65		0.45
Cd	0.0002		0.0001
Cr	0.28	84	0.28
Cr(VI)	0.006	-	0.009
Cu	0.006	3,700	0.003
Fe	<0.05	5,770	<0.05
Ga	<0.5	5,200	<0.5
Hg	<0.0002	-	<0.0002
I	<1	-	<1
Mn	<0.03	26	<0.03
Mo	2.1		2.1
Ni	0.005	63	<0.005
Pb	<0.002		<0.002
Se	0.030	410	0.027
Sr	6.9	710	6.9
Ti	<1	-	<1
U	<0.02	2.1	<0.02
V	<0.5	1500	<0.5
Zn	<0.02	146	<0.02
TDS (105°C)	2540	$1x10^6$	2490
Total Sol. SiO_2	3.7	-	1.8
Hardness (as $CaCO_3$)	1450	-	1450

2.1.3 Operational Details and Results of 'Seeded' RO

'Seeded' RO treatment of neutralized fly ash leachate (pH 7.3) was carried out at 5.5 MPa applied pressure at a recirculatory feed flow rate of ~19 L/min. During operation, feed was added from storage to the batch holding tank (target level maintained at ~40 L) at intervals as required to make up for loss of liquid as permeate was removed from the system ('feed-and-bleed' mode). At the end of each day's operation, the level of RO feed/retentate in the holding tank was allowed to drop to near 'empty', to check whether the attainable limit of volume reduction was being reached. The RO lines were pumped out and given a brief flush (~5 s) with deionized water to prevent possible overnight membrane blinding by precipitation of calcium sulphate from supersaturated solution.

Throughout the run, which lasted for a total of 2 153 min (~36 h), flux rates, pH, and specific conductivity (a rapid, conveniently measured indicator of TDS levels) were monitored for individual Tubes A and B. These observations are summarized in Table II.

TABLE II

'SEEDED' RO OF FRESH LAKEVIEW TGS FLY ASH LEACHATE AT 5.5 MPA APPLIED PRESSURE
'ABCOR' TUBULAR MEMBRANE TEST PAIR: 'FEED-AND-BLEED' BATCH MODE

STREAM	PARAMETER	MINIMUM	MEAN	MAXIMUM
Feed/Retentate	pH	7.3	7.8	8.1
	Sp Cond (mS/m)	260	525	1015
Permeate (Tube A)	pH	6.8	7.4	7.8
	Sp Cond (mS/m)	6	12	49
	Flux Rate (mL/min)	51.3	74.5	92.5
	Rejn (Cond)*(%)	95.3	97.0	98.4
Permeate (Tube B)	pH	6.9	7.4	7.8
	Sp Cond (mS/m)	6	12	49
	Flux Rate (mL/min)	60.0	77.1	95.0
	Rejn (Cond)*(%)	95.2	97.1	98.1

* Conductivity rejection efficiency is calculated as $(1-C_p/C_f) \times 100$ where C_p is the instantaneous specific conductivity of the permeate and C_f is the corresponding specific conductivity of the feed/retentate

Membrane Tubes A and B gave closely matching RO flux and specific conductivity rejection characteristics. Figure 2 illustrates flux/rejection characteristics versus time for the test pair; a linear regression analysis of flux versus time showed a flux decline of 0.00123 mL/min for the assembly.

At the end of the test run, 338 L of neutralized leachate had yielded 17.5 L of concentrate and 320.5 L of permeate - corresponding to an overall volume reduction factor of 20:1, or 95% conversion to permeate. This is not

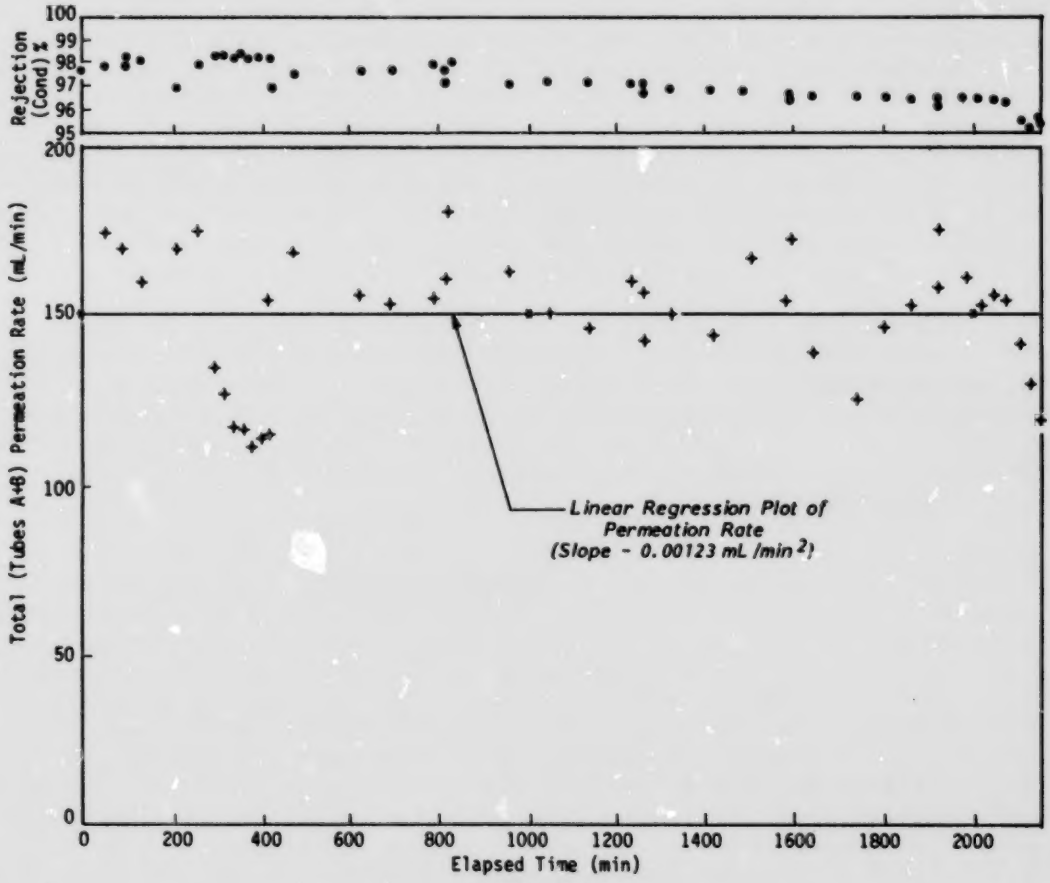

FIGURE 2
FLUX AND REJECTION CHARACTERISTICS VS TIME FOR
'SEEDED' RO TEST WITH FLY ASH LEACHATE

considered to be the maximum attainable, since the final retentate volume was fixed by the hold-up volume of the system (centrifuge, plumbing, etc).

Standard RO flux/rejection tests were carried out before and after the 'seeded' RO run, using NaCl(aq) at 5 000 mg/L. There was no evidence of significant membrane fouling. Volume reduction of fly ash leachate is therefore eventually likely to reach a practical limit when the concentrations of soluble sodium, potassium and sulphate ions in the retentate give rise to a combined osmotic pressure effectively opposing the applied pressure. (The concentration of calcium sulphate is limited by its solubility limit of about 2 000 mg/L as $CaSO_4$).

About 650 g of gypsum solids were recovered from the centrifuge. Samples of final retentate, composite permeate and centrifuge solids were submitted for detailed chemical analysis. The results are given in Table III. Mass balance calculations across the 'seeded' RO system show that Ca, Mg, K, B, Sr and TDS were >90% accounted for. About 80-90% of SO_4=, Cl^- and Na were accounted for. For carbonate a very poor mass balance was observed. The centrifuge solids contained a very high level of carbonate, possibly formed in the system by absorption of atmospheric CO_2 into the retentate (pH 7-8) which was continuously recirculated during the 'seeded' RO test run. The quality of 'seeded' RO product water (permeate) was found to be extremely good, and complies throughout with published Ontario MOE Objectives for surface waters/4/.

Where adequate data is available, calculated RO rejection efficiencies for most solutes (Table III) were found to greatly exceed 90%, with the notable exception of boron, which was rejected by only 39.4%. This is not unexpected, however, since boron rejection is well-known to be poor when the element is present as boric acid, which is feebly ionized in neutral or acidic aqueous solution, and passes easily through RO membranes by virtue of its hydrogen-bonding properties.

2.2 RO FOR BORON REMOVAL

It was of interest to test the feasibility of applying RO for further boron removal (if required), utilizing the fact that boric acid can be readily ionized to borate by addition of alkali (eg NaOH). Since borate was expected to be relatively well rejected by RO membranes, a separate RO test was carried out on the permeate (containing 20 mg/L boron) from the 'seeded' RO study. A DuPont B-9 'Permasep' polyamide hollow fibre permeator was used, since cellulose acetate membranes are susceptible to hydrolysis and deteriorate rapidly at elevated pH. The B-9 module, pre-tested with a standard 5 000 mg/L NaCl(aq) solution, at 2.1 MPa, had a NaCl rejection efficiency of 92% with a flux rate of 7.92 m^3/module•d.

A 75-L batch of permeate from the 'seeded' RO test was adjusted from pH 6.7 to pH 11.0 by addition of NaOH. This was then used as feed to the B-9 module.

The RO test using the B-9 'Permasep' permeator for boron removal lasted only about 12 min, during which time maximum pacticable conversion to permeate took place, limited only by volumetric holdup in the system. Results of the test are summarized in Table IV.

Clearly, improved boron removal from fly ash leachate (99+%), if required, can be readily achieved by RO operation at elevated pH.

The permeate, at about pH 10.7 was easily re-adjusted to within the MOE Objective range of pH 6.5 - 8.5 by minor addition of H_2SO_4.

3. ENVIRONMENTAL IMPLICATIONS

The major ion chemistry of the various 'seeded' RO process streams is summarized in the Piper diagram (Figure 3), which contains also data for surface waters sampled upstream of the ash disposal area. The Piper diagram shows that the RO feed (before and after neutralization) has dominance of alkaline earths and strong acid anions, while both the RO concentrate and permeate have dominance of alkali metal cations and strong acid anions.

TABLE III

ANALYTICAL DATA FOR 'SEEDED' RO OF FRESH LAKEVIEW FLY ASH LEACHATE AT 5.5 MPa APPLIED PRESSURE

Quantity	338 L	17.5 L	320.5 L		0.650 kg		Mass Balance Across 'Seeded' RO System (where sufficient data available)		
Stream	RO Feed After Neutralization (mg/L)	Final RO Concentrate (mg/L)	Cumulative RO Permeate (mg/L)	MOE Water Quality Objectives (1978) (mg/L)	Solids After Centrifugation (mg/kg)	RO Rejection Efficiency* (%)	Σ (Input) (mg)	Σ (Output) (mg)	Recovery (%)
pH	7.3	8.0	6.7	6.5-8.5		98.1			
Sp. Cond (mS/m)	260	971	12	–		–	–	–	
Alkalinity OH^-	<5	<5	<5	Not to decrease by more than 25% of natural concentration	135,000	–	–	87,750 (solids)	N.D.**
as CO_3	<5	<5	<5		–	92.0	9,126 (aq)	4,076 (aq)	
as HCO_3^-	27	123	6		612,000	98.3	574,600	491,500	86
$CaCO_3$	2700	4420	51	–	70	85.7	3,143	2,482	79
SO_4	9.3	55	4.6	–	–				
Cl	–	<25	0.51	–	270,000	98.5	196,040	187,600	96
NO_3	560	540	8.2	–	–	99.4	112	113	101
Ca	0.33	6.1	0.02	–	–	98.2	28,730	25,200	88
Mg	85	1220	12	–	1,100	98.4	19,942	18,070	91
Na	59	860	7.2	–	–				
K	<0.0002	0.0006	<0.0002	<0.0001	585				
Ag	<0.5	<0.5	<0.5	Undefined Tolerance Limit	2				
Al	0.02	<0.0002	<0.002	<0.1	–	39.4	7,436	7,180	97
As	22	44	20	Undefined Tolerance Limit	–				
B	<0.0002	<0.0002	<0.0002	0.11 (Hardness <75 mg/L)					
Ba	<1	<1	0.37	Undefined Tolerance Limit	150				
Be	0.45	2.7	0.0002†	–	–				
Br	0.0001	0.01	<0.05	<0.0002	57	96.0			
Cd	0.28	1.0	0.009	<0.1		92.2+			
Cr	0.009	1.0	0.005†	–	–	98.2			
Cr(VI)	0.003	0.08	<0.05	<0.005	1,610	88.0			
Cu	<0.05	<0.05	<0.5	<0.3	18				
Fe	<0.5	<0.5	<0.0002	–	–				
Ga	<0.0002	<0.0002	0.044	<0.0002	–				
Hg	<1	0.15	<0.03	–					
I	<0.03	<0.03	<0.5	Undefined Tolerance Limit	7				
Mn	2.1	28	<0.005	Undefined Tolerance Limit	–	96.7+			
Mo	<0.005	2.4	<0.002	<0.025	380				
Ni	<0.002	0.002	0.008	<0.025 (Alk. >80 mg/L)	–				
Pb	0.027	0.91	0.17	<0.1	16	98.3			
Se	6.9	30	<1	–	3,000	99.1	2,332	2,530	108
Sr	<1	<1	<0.02	–	–				
Ti	<0.02	<0.02	<0.5	–	1.7				
U	<0.5	<0.05	<0.02	Undefined Tolerance Limit	21				
V	<0.02	0.04	157	<0.03	58				
Zn	2490	8000	0.6	Increase <1/3 background	$1x10^6$	96.6	841,620	840,320	100
TDS (105°C)	1.8	11.3	20.6	–	–	90.8			
Total Sol. SiO_2	1450	1375		–	–	–			
Hardness (as $CaCO_3$)									

* RO Rejection Efficiency = $(1 - C_p/C_{av}) \times 100$ where C_p = concentration of permeate and C_{av} = arithmetic mean of concentrations of feed and concentrate.

** N.D. denotes not determined - due to excessively high carbonate levels found, attributable to absorption of atmospheric CO_2 by recirculating retentate at pH 7-8

†Elevated levels considered to be due to analytical uncertainties at these low concentrations.

Comparison Data Points

- ○ *Before Ash Placement Commenced*
- ● *After Ash Placement Commenced*
- ▲ RO *Feed After Neutralization*
- ▼ RO *Feed Before Neutralization*
- □ RO *Permeate*
- ◇ RO *Concentrate*

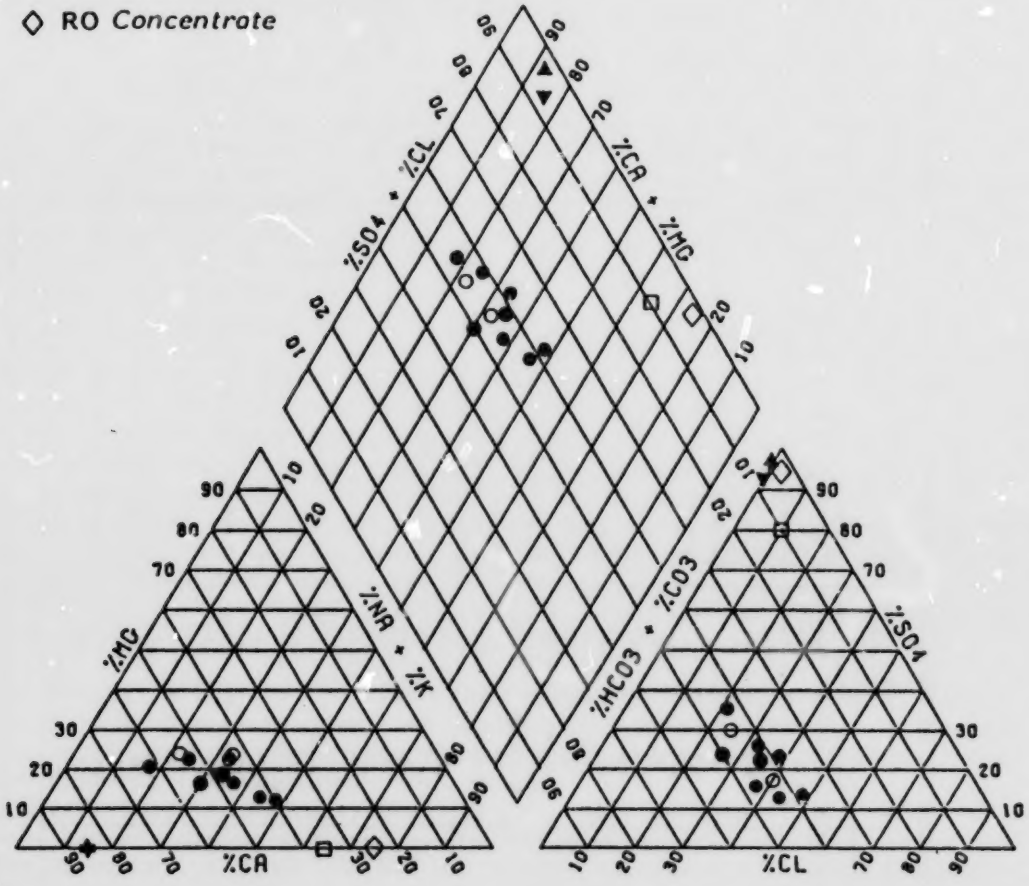

FIGURE 3

'SEEDED' RO PROCESS STREAMS:
PIPER DIAGRAM SHOWING MAJOR ION CHEMISTRY
(COMPARISON LOCATION: DOMTAR QUARRY,
MARY FIX CREEK, CP RAILWAY OVERPASS)

TABLE IV

BORON REMOVAL BY RO USING DUPONT HOLLOW FIBRE B-9 'PERMASEP' PERMEATOR AT 2.8 MPa USING PERMEATE (ADJUSTED TO pH 11.0) FROM 'SEEDED' RO TEST AS FEED

	Feed	Retentate	Permeate	Rejection† (%)	Flux Rate* (m^3/module·d)
Volume (L)	75.0**	4.5**	70.5	-	
pH	11.0	11.7	10.7	-	
Sp Conductivity (mS/m)	114	602	35	90.2	6.4
B (mg/L)	20	180	0.9	99.1	

† Rejection efficiency is calculated here as $(1-Cp/C_{av}) \times 100$ where C_{av} is the arithmetic mean of initial feed and final retentate concentrations, and C_p is the permeate concentration.

* Membrane surface area of B-9 not known; flux rate quoted per module.

** Incomplete recovery of boron in retentate due to B-9 test loop system holdup - estimated to be ~3 L, corresponding to a boron recovery of ~95%.

Using cellulose acetate membrane modules, RO rejection efficiency was very high for all species except boron. If required, boron may be effectively removed from the permeate by including a secondary RO treatment stage, using polyamide hollow fibre modules.

As pointed out earlier, so far there has been no need for treatment of the leachate, and the 'seeded' RO treatment is strictly a contingency plan. The 'seeded' RO concentrate may be used for spraying on-site as a fly ash dust suppressant, or may be trucked away to a liquid waste disposal site, as appropriate. The solid materials produced in the course of 'seeded' RO of fly ash leachate (ie the precipitate formed after pH adjustment and the gypsum solids removed by centrifugation) are not hazardous, and may either be disposed of by on-site placement together with incoming fly ash, or else trucked off to an industrial landfill waste disposal site.

4. COST ESTIMATES

Based on operational data generated in the course of the 'seeded' RO test, costs have been estimated (Table V) for RO treatment of fly ash leachate at three levels of operation. The principal assumptions made are:

(i) In estimating equipment size and cost, a flux rate of 0.81 $m^3/m^2 \cdot d$ at 5.5 MPa applied pressure applies.

TABLE V

SUMMARY OF PROJECTED COST ESTIMATES FOR FLY ASH LEACHATE TREATMENT BY 'SEEDED' REVERSE OSMOSIS (RO)
Basis: Usage 240 days/a @ 24 h/d; RO Flux rate = 0.81 m^3/m^2 d at 5.5 MPa; Volume reduction factor = 20

Leachate Volume (m^3/a)	910*	10,900**		54,500***	
Treatment System Type	Stationary	Stationary	Mobile	Stationary	Mobile
RO Assembly					
RO Unit Nominal Capacity (m^3/d)	3.785	41.25	41.25	206.3	206.3
Capital Cost (Incl Fed & Prov Sales Tax & Duty)	$17,000.00	$170,000.00	$170,000.00	$425,000.00	$425,000.00
Installation Cost	$ 3,400.00	$ 34,000.00	$ 34,000.00	$ 42,000.00	$ 42,000.00
Total Installed Cost of RO Assembly	$20,400.00	$204,000.00	$204,000.00	$467,000.00	$467,000.00
Amortized Capital Cost of RO/a	$ 3,050.00	$ 30,500.00	$ 30,500.00	$ 69,700.00	$ 69,700.00
Amortized Capital Cost of RO/m^3 Permeate	$ 3.35	$ 2.80	$ 2.80	$ 1.28	$ 1.28
Operating & Maintenance Cost of RO/m^3 Permeate†	$ 8.98[a]	$ 1.44[a]	$ 1.44[a]	$ 0.74[a]	$ 0.74[a]
Total RO Cost/m^3 Permeate	$ 12.33	$ 4.24	$ 4.24	$ 2.02	$ 2.02
Centrifuge System					
Centrifuge Loading Capacity (L/min)	0.13	1.6	1.6	7.9	7.9
Capital Cost (Incl Fed & Prov Sales Tax & Duty)	$16,000.00	$ 29,000.00	$ 29,000.00	$ 42,000.00	$ 42,000.00
Installation Cost	$ 2,000.00	$ 4,000.00	$ 4,000.00	$ 8,000.00	$ 8,000.00
Total Installed Cost of Centrifuge System	$18,000.00	$ 33,000.00	$ 33,000.00	$ 50,000.00	$ 50,000.00
Amortized Capital Cost of Centrifuge/a	$ 2,700.00	$ 4,900.00	$ 4,900.00	$ 7,500.00	$ 7,500.00
Amortized Capital Cost of Centrifuge/m^3 Permeate	$ 2.97	$ 0.45	$ 0.45	$ 0.14	$ 0.14
Operating & Maintenance Cost of Centrifuge/m^3 Permeate†	$ 8.16[a]	$ 1.36[b]	$ 1.36[b]	$ 0.54[c]	$ 0.54[c]
Total Centrifuge Cost/m^3 Permeate	$ 11.13	$ 1.81	$ 1.81	$ 0.68	$ 0.82
Trucking (for Mobile Treatment System)					
Truck Rental Charges/a (240 days @ $175/wk)			$ 6,000.00		$ 6,000.00
Distance Surcharges/a (4000 km @ $0.12/km)			$ 500.00		$ 500.00
Fuel & Maintenance/a			$ 3,800.00		$ 3,800.00
Labour Charges/a			$ 10,000.00		$ 10,000.00
Total Trucking Cost/m^3 Permeate			$ 1.86		$ 0.37
Building (for Stationary Treatment System)					
Capital Cost of Building	$ 9,600.00	$ 57,700.00		$135,400.00	
Amortized Capital Cost of Building/a	$ 1,400.00	$ 8,600.00		$ 23,800.00	
Building Cost/m^3 Permeate	$ 1.54	$ 0.79		$ 0.44	
TOTAL COSTS/m^3 Permeate	$ 25.00	$ 6.84	$ 7.91	$ 3.07	$ 3.07
TOTAL COSTS/a	$22,800.00	$ 74,600.00	$ 86,200.00	$167,300.00	$167,300.00

* Equivalent to about 1 sump-full/a of Domtar fly ash leachate
** Equivalent to about 1 sump-full/month of Domtar fly ash leachate
*** Equivalent to about 1 sump-full/week of Domtar fly ash leachate

or comparable volumes to be treated at other locations in Southern Ontario

† It is assumed that operating and maintenance labour at $30/h is required as follows: a = 1 h/d; b = 2 h/d; c = 4 h/d.

(ii) Capital and operating costs are based on current information provided by equipment suppliers; amortization rates were calculated over 20 years at a prime rate of 12%, using methodology published by the U.S. DOE/EPA /5/.

(iii) Continuous treatment of fly ash leachate is assumed, corresponding to:

(a) one sump-volume of Domtar leachate per year (910 m^3/a)

(b) one sump-volume of Domtar leachate per month (10,900 m^3/a)

and (c) one sump-volume of Domtar leachate per week (54,500 m^3/a).

(iv) Should fly ash leachate treatment be required at a number of different sites, a mobile 'seeded' RO treatment system is justified.

For the largest-scale 'seeded' RO operation considered, ($3.07/$m^3$), the corresponding costs of evaporation are estimated/5/ to be $4.05/$m^3$. On this basis, costs of 'seeded' RO are about 75% of the costs of evaporation per cubic metre of purified leachate.

Should more complete boron removal from the 'seeded' RO permeate be required, the additional costs for secondary RO treatment are as estimated in Table VI for each level of operation. For the largest scale, total RO treatment costs would then become $4.79/$m^3$, ie about 18% higher than for evaporation.

TABLE VI

COSTS OF BORON REMOVAL BY RO USING B-9 'PERMASEP' PERMEATOR AT 2.8 MPA APPLIED PRESSURE

Leachate Volume (m^3/a)	910	10,900	54,500
B-9 Nominal Capacity (m^3/module·d)	3.785	41.25	206.30
Capital Cost	$13,000	$130,000	$325,000
Installation Cost	$ 2,600	$26,000	$32,500
Total Installed Cost	$15,600	$156,000	$357,500
Amortized Capital Cost/a	$ 2,300	$23,300	$53,400
Operation & Maintenance Cost/m^3 of permeate*	$2.53	$2.14	$0.98
Total B-9 Cost/m^3 permeate	$11.51	$3.58	$1.72
Total B-9 Cost/a	$10,500	$39,000	$93,700

* It is assumed that 1 h/d of operating and maintenance labour is required at $30/h.

There exists also the possibility that 'seeded' RO may be found applicable to treatment of other gypsum-laden waste waters besides fly ash leachate, eg where lime has been used to neutralize acidic waste streams containing sulphuric acid (such as coal pile drainage) and where impurities may endanger water quality at the point of discharge. In such instances a mobile 'seeded' RO system may be further justified, with maximum equipment utilization and economy.

5. CONCLUSIONS

1. 'Seeded' RO treatment of fly ash leachate was found to be technically feasible, based on the results of a study using a tubular cellulose acetate RO membrane test rig in conjunction with a triple bowl centrifuge.

2. RO membrane fouling was effectively avoided: gypsum precipitated from saturated solution as finely suspended nuclei that grew to a size removable by in-line centrifugation.

3. Dissolved components of fly ash leachate were effectively removed by RO membrane separation, yielding a product water (permeate) of extremely high quality, the constituents of which were all below published Ontario MOE Objectives for surface waters.

4. A minimum volume reduction factor for fly ash leachate of about 20:1 was found to be attainable; this was experimentally limited by liquid holdup in the test rig.

5. Relatively poor boron rejection efficiency observed can be greatly improved, if deemed necessary from an environmental standpoint, by secondary processing of the permeate at elevated pH using a polyamide hollow fibre permeator, re-adjusting the pH down to within acceptable limits prior to discharge.

6. Solid by-products generated during pretreatment and operation may be either disposed of on-site as landfill or hauled to an off-site disposal facility.

7. The retentate may be used for on-site fly ash dust suppression, or hauled off-site to an industrial liquid waste treatment facility.

8. Preliminary cost estimates for 'seeded' RO treatment, based as appropriate on stationary and mobile treatment systems, have been made at three different levels of operation. The economy of scale is clearly evident, with projected total costs ranging from about \$3 to \$25/m^3 of permeate, depending on scale and mode of operation. 'Seeded' RO costs are estimated to be less than 75% of the costs of evaporation per cubic metre of purified leachate for the largest scale considered.

9. Secondary RO treatment for more complete boron removal (if required) is estimated to bring the total RO costs into the range \$5 to \$37/m^3 of permeate, depending on scale. Total RO costs would then be about 18% higher than for evaporation per cubic metre of purified leachate for the largest scale considered.

ACKNOWLEDGEMENT

The author wishes to thank Mr. A.Z. Mlynarczyk for his able technical assistance in performing the experimental work.

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ROLE DE L'ACTIVITE BIOLOGIQUE DANS UN FILTRE A SABLE CONVENTIONNEL

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RESUME

La filtration sur lit de sable se répand de plus en plus comme technique de polissage des effluents d'usine d'épuration. La biomasse qui se développe inévitablement dans les lits filtrants et en accélère le colmatage est habituellement considérée comme une nuisance. La présente étude a été entreprise pour évaluer l'efficacité de l'activité biologique dans un filtre à sable pour l'enlèvement de la DBO soluble; et pour préciser la corrélation existant entre la biomasse et les autres paramètres de la filtration tels: la perte de charge, la consommation d'oxygène, l'enlèvement de la DBO_5 et des nitrates.

Le montage expérimental consistait en une colonne de filtration de 133 mm de diamètre. Le milieu filtrant était constitué de 26 cm de sable d'Ottawa ayant un diamètre effectif de 1,30 mm et un coefficient d'uniformité de 1,50. L'eau usée synthétique, à base de sucrose, contenait environ 35 mg/L de DBO_5 soluble et 10 mg/L de nitrates. Des essais de filtration ont été effectués à des taux de 71,6 et 143 L/m^2-min (1,8 et 3,5 g/pi^2-min) et les résultats montrent que l'on peut atteindre un enlèvement de 40 à 50% de la DBO_5.

Un cycle de filtration biologique se déroule en trois (3) phases similaires à celles de la croissance bactérienne: acclimatation, croissance rapide et plafonnement. L'enlèvement des nitrates n'est significatif que dans la dernière phase. Il y a une excellente corrélation entre la DBO_5 et le carbone organique ainsi qu'entre l'enlèvement du carbone et la consommation d'oxygène. Il y a une bonne corrélation linéaire entre l'enlèvement de carbone et le poids de la biomasse dans le filtre dans la seconde phase de la filtration. Au début et à la fin du cycle, la corrélation est faible.

La perte de charge est une fonction de puissance du poids de la biomasse dans les deux (2) premières phases de la filtration. Dans la dernière phase, cette relation est complètement modifiée parce que la densité de la biomasse diminue très rapidement.

INTRODUCTION

A mesure que les normes de rejet dans la nature deviennent plus contraignantes, la nécessité se fait souvent sentir de traiter encore

davantage les effluents des usines d'épuration dont la qualité, hier encore, était jugée satisfaisante; et les filtres à sable classiques, à cause de leur simplicité, sont souvent considérés comme une méthode intéressante d'effectuer cette opération de polissage.

La croissance àpeu près inévitable de matière biologique dans les filtres à sable est habituellement considérée comme une nuisance et on s'efforce de la prévenir ou de l'éliminer.

Notre monde tend cependant de plus en plus vers la conservation et le recyclage, vers l'utilisation des énergies douces et des moyens biologiques pour différentes fins. Un peu dans ce courant, nous nous sommes demandés si l'on ne pourrait pas utiliser plutôt que détruire cette biomasse des filtres à sable.

Cette étude n'est certes pas la première à être effectuée sur ce phénomène. Des chercheurs de l'Université des Sciences et des Techniques du Languedoc à Montpellier, en France, Messieurs Grasmick et al. (1, 2, 3) ont fait des études théoriques et expérimentales poussées sur la filtration biologique immergée avec des matériaux granulaires poreux.

Il nous est apparu intéressant d'évaluer l'efficacité que peut avoir l'activité biologique dans un filtre à sable conventionnel pour l'enlèvement de la DBO soluble et des nitrates; et de préciser la corrélation existant entre la biomasse et les autres paramètres de la filtration tels: la perte de charge, la consommation d'oxygène, l'enlèvement de la DBO_5 et des nitrates. L'implantation de la biomasse, sa résistance au lavage, sa vitesse d'acclimatation à des variations brusques de concentration en substrat faisaient également partie des objectifs de la recherche.

APPAREILLAGE D'ESSAIS

Comme on peut le voir à la figure 1, l'élément central du montage d'essai est constitué d'une colonne en plexiglas de 133 mm de diamètre et de près de 2 mètres de hauteur. Elle est équipée de tous les accessoires et instruments nécessaires au contrôle et à la mesure du taux de filtration, du débit d'eau et d'air du lavage ainsi que de l'aération et de l'alimentation en substrat. Elle est également équipée d'une série de prises pour l'échantillonnage liquide et solide à différents niveaux ainsi qu'à la mesure des pressions hydrauliques en différents points.

Une charge hydraulique constante est maintenue sur le filtre grâce à un déversoir de trop-plein au sommet de la colonne.

Le milieu filtrant est constitué de 26 cm de sable d'Ottawa ayant un diamètre effectif de 1,30 mm et un coefficient d'uniformité de 1,50. Le sable est séparé des agrégats sous-jacents par un tamis aux mailles suffisamment fines pour empêcher tout entrainement lors de la filtration.

L'eau usée synthétique utilisée pour les essais (tableau 1) est une solution de sucrose à laquelle sont ajoutés différents sels minéraux nécessaires à l'activité biologique. Cette recette a été utilisée avec satisfaction dans d'autres études sur l'épuration biologique (4). On y ajoute du nitrate de potassium pour évaluer l'enlèvement des nitrates.

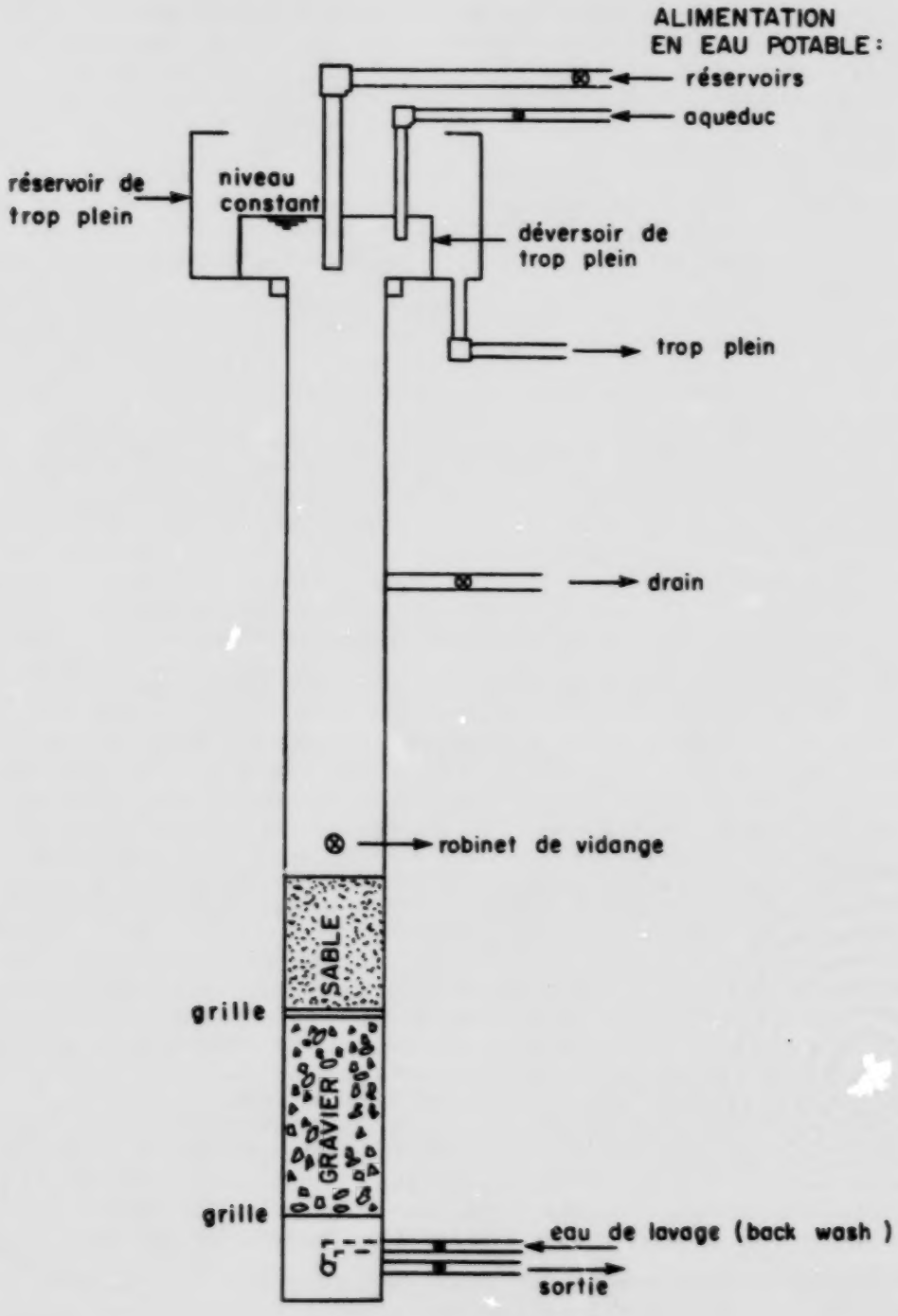

Figure 1. Croquis de la colonne de filtration

TABLEAU 1
FILTRATION BIOLOGIQUE
EAU USEE SYNTHETIQUE

CONSTITUANTS	CONCENTRATION (mg/L)
Sucrose	50,0
$(NH_4)_2SO_4$	25,0
K_2HPO_4	53,5
KH_2PO_4	26,3
$MgSO_4 \cdot 7\ H_2O$	10,0
$MNSO_4 \cdot H_2O$	1,0
$FeCl_3 \cdot 6\ H_2O$	0,05
$CaCl_2$	0,75
KNO_3	70,0

BREVE DESCRIPTION DE L'ENSEMENCEMENT ET DES ESSAIS

L'utilisation d'un filtre pour la filtration biologique suppose l'existence dans le filtre d'une biomasse bien implantée, uniformément répartie, et acclimatée aux conditions d'opération: type et concentration du substrat ainsi que taux de filtration. Une telle implantation peut prendre de deux (2) à quatre (4) semaines.

La figure 2 montre, à titre d'exemple, l'ensemencement du filtre préparatoire aux essais de filtration. On commence par effectuer une recirculation d'eau usée municipale ou de boue activée diluée. Cette eau doit être bien aérée et changée périodiquement. Après quelques jours, on ajoute à l'eau de recirculation une faible quantité du substrat choisi pour permettre aux microorganismes de s'acclimater à ce substrat.

Lorsque la biomasse commence à être bien implantée (ce qui se manifeste par une consommation plus importante d'oxygène dissous), on cesse la recirculation pour ne faire circuler dans le filtre qu'une solution du substrat choisi. La concentration du substrat est augmentée graduellement jusqu'à la valeur choisie pour l'essai de filtration.

Chaque fois que la perte de charge atteint une valeur limite, on procède au lavage du filtre et on recommence la filtration.

Trois (3) essais ont été effectués, dont les principales caractéristiques sont montrées au tableau 2. Un premier essai, qu'on pourrait qualifier d'avant-première, a plutôt servi à préparer le déroulement de l'essai suivant. On y a prélevé cinq (5) échantillons à intervalles irréguliers.

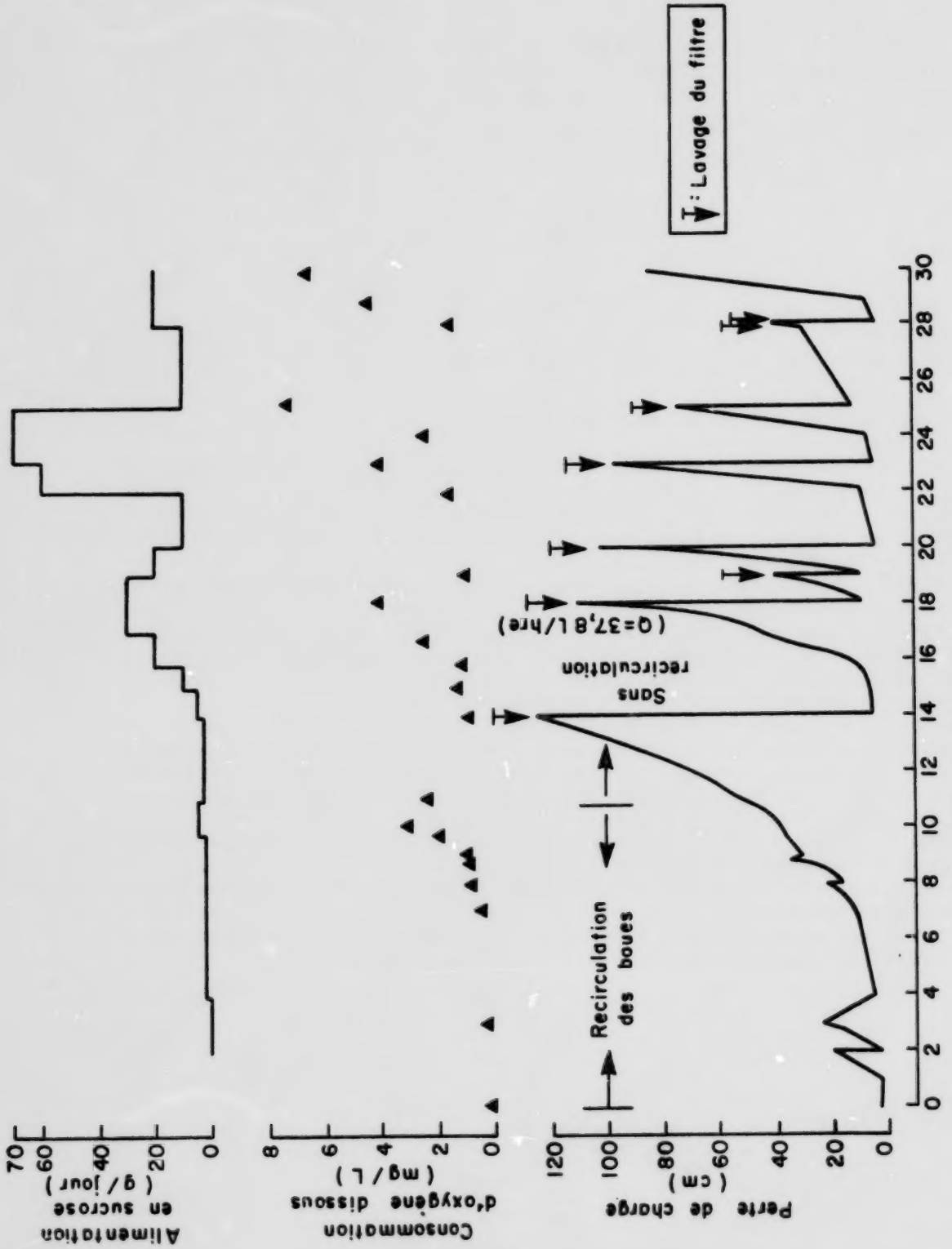

Figure 2. Ensemencement du filtre

TABLEAU 2
CARACTERISTIQUES
DES TROIS ESSAIS DE FILTRATION

ESSAI	DUREE (HEURES)	TEMPS ENTRE LES ECHANTILLONS	NOMBRE D'ECHAN-TILLONS	CONDITIONNEMENT A LA CONCENTR. EN SUBSTRAT	TAUX FILTRATION (L/m^2-min.)
1	48	Irrégulier	5	Oui	71,6
2	52	2 heures	27	Non	71,6
3	20	1-2 heures	13	Oui	143,0

Le deuxième essai a eu lieu quelques jours plus tard au même taux de filtration de 71,6 L/m^2-min. Les vingt-sept (27) échantillons prélevés à intervalles de 2 heures permettent de suivre à la trace l'évolution des différents paramètres et d'effectuer une bonne analyse de la corrélation qui existe entre eux. L'alimentation en substrat a été faite à taux réduit dans l'intervalle de temps séparant les deux (2) essais.

Le troisième essai a été fait un (1) mois plus tard à un taux de filtration double de celui des deux (2) précédents.

RESULTATS

Avant de présenter la performance globale de ce mode de filtration, il est intéressant de jeter un coup d'oeil sur l'évolution des différents paramètres et sur la corrélation qui existe entre eux.

a) Evolution des paramètres

D'une façon générale, les paramètres de la filtration évoluent en suivant les trois (3) phases de la croissance bactérienne: adaptation, croissance et plafonnement ou stagnation. On pourrait dire que le phénomène suit la courbe en S classique de l'évolution de la population.

Biomasse

Cette évolution est très évidente si on prend l'évolution du poids unitaire de la biomasse dans le filtre (figure 3). On voit une phase d'acclimatation d'une quinzaine d'heures, une phase de croissance d'une vingtaine d'heures et une phase de stagnation et même de décroissance d'une quinzaine d'heures.

En dépit d'un certain flottement, dû aux limites de l'échantillonnage et à la nature même du phénomène, on pourrait tracer une courbe en S qui ajusterait très bien les résultats. La dispersion des points à la fin du cycle s'explique

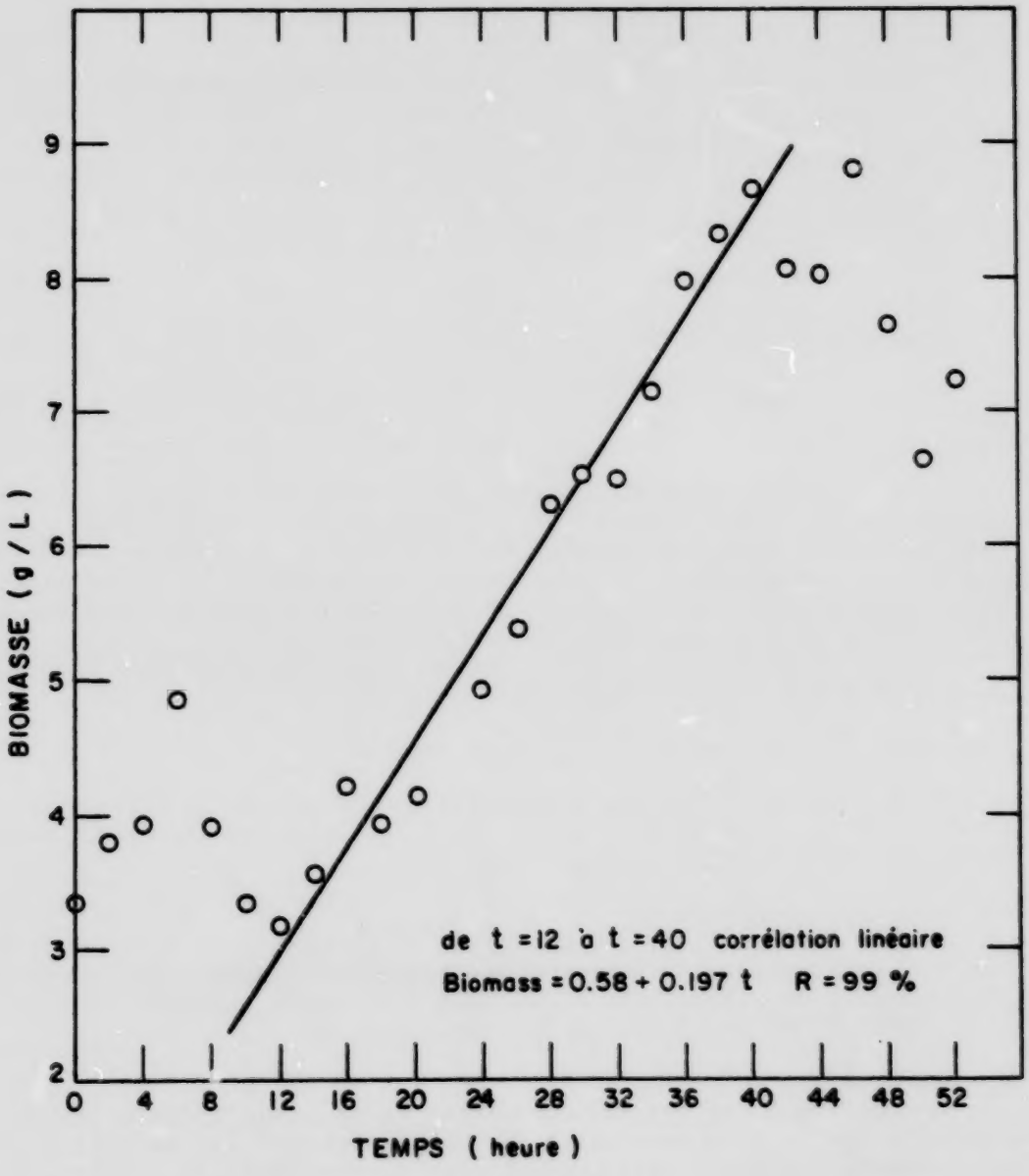

Figure 3. Deuxième essai: évolution de la biomasse

en bonne partie par la desquamation (ou décrochage) sporadique de la matière organique parvenue à maturité.

Oxygène dissous

On voit que l'évolution de la consommation d'oxygène dissous suit une très belle courbe en S (figure 4). Un ajustement de la courbe par régression polynominale a également été fait parce que c'est une formulation mathématique qui s'utilise plus commodément dans un modèle.

Enlèvement du carbone et de la DBO_5

En ce qui concerne l'épuration, disons d'abord que le carbone organique total et la DBO_5 ont été mesurés sur une série d'échantillons ayant entre 4 et 50 mg/L de DBO_5 et que la corrélation est très bonne pour le substrat utilisé, comme on peut le voir par la figure 5.

Le carbone organique total a été mesuré avec un analyseur de carbone DOHRMAN, DC-80 avec échantillonneur automatique et la DBO_5 selon la méthode standard par mesure de l'oxygène dissous avec un oxymètre avec sonde à membrane. Comme la mesure du carbone organique total est très rapide avec un tel appareil, seul celui-ci a été mesuré par la suite. Mais tout ce qui est dit du carbone organique vaut évidemment pour la DBO_5.

En dépit de l'allure un peu moins bien définie de l'évolution de l'enlèvement du carbone (figure 6), les trois (3) phases d'adaptation de croissance et de plafonnement y sont assez évidentes.

Enlèvement des nitrates

La teneur en nitrates a été mesurée avec un autoanalyseur avec échantillonneur automatique, un autre appareil très performant.

L'évolution de l'enlèvement des nitrates fait exception au patron général qui a été présenté (figure 7). A peu près nul durant plus de la moitié du cycle, l'enlèvement des nitrates démarre brusquement et se maintient à environ 1 mg/L jusqu'à la toute fin alors le filtre est très colmaté. A ce moment, on remarque une forte augmentation de l'enlèvement. On verra plus loin l'explication de ce phénomène en parlant des corrélations entre les différents paramètres.

Perte de charge

La perte de charge échappe également au patron en S que suivent les autres paramètres. Elle augmente exponentiellement avec le temps (figure 8). Le léger lavage effectué à la 38e heure pour faciliter la prise d'échantillons solides n'a fait, en dilatant le filtre, que décaler la courbe d'environ cinq (5) heures. La ligne pointillée représente l'évolution présumée sans ce délai.

Si on regarde la perte de charge dans les cinq (5) tranches du filtre (figure 9), on voit qu'au début, la perte charge unitaire est huit (8) fois plus élevée dans la tranche supérieure que dans les quatre (4) autres alors qu'à la fin, il y a un gradient bien marqué entre la perte de charge de chaque tranche et celle du dessus n'est que trois (3) fois celle du dessous.

b) Corrélation entre les paramètres

Voyons maintenant la corrélation entre les différents paramètres. Le but de la recherche étant d'évaluer les possibilités du filtre à sable gravitaire

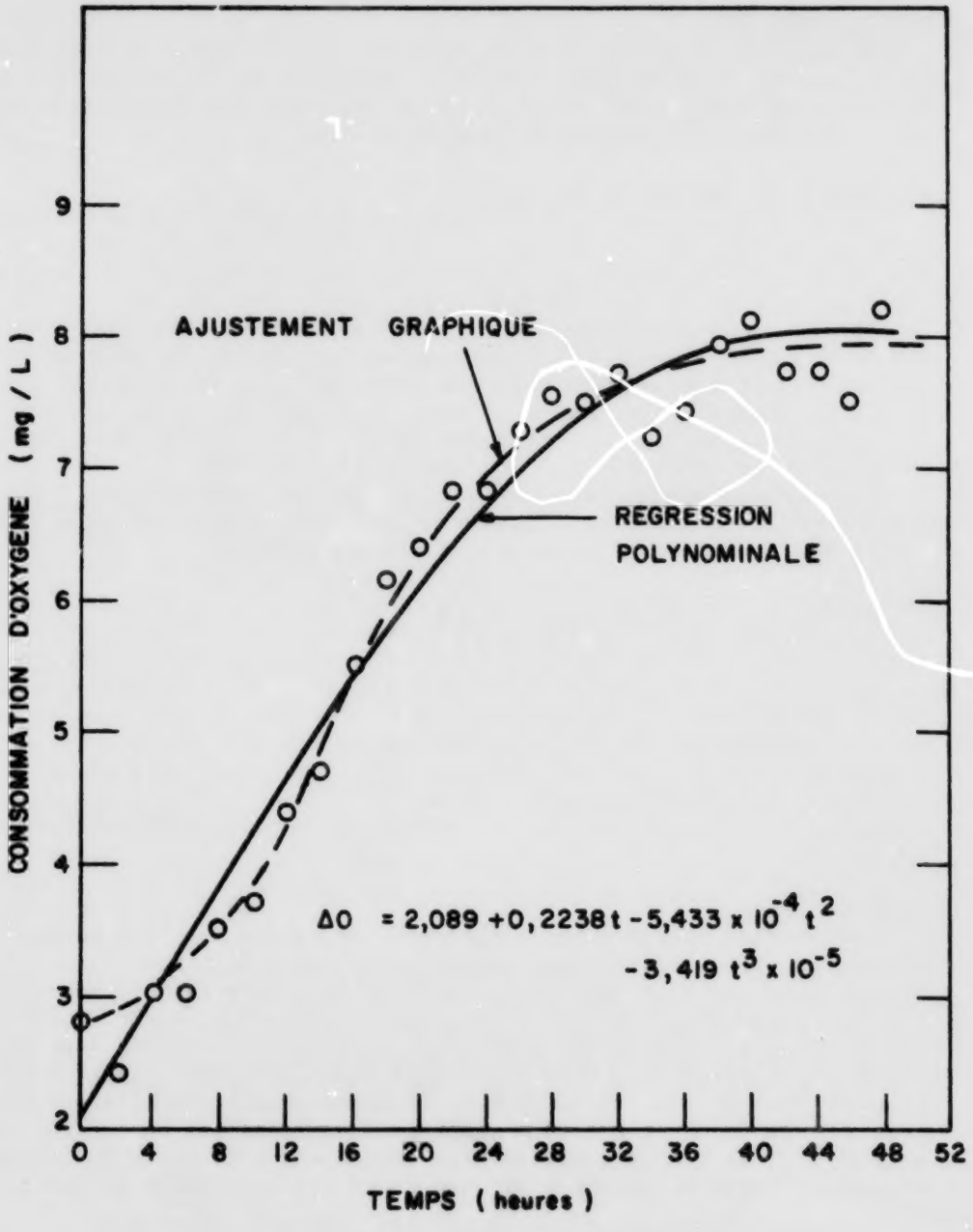

Figure 4. Deuxième essai: évolution de la consommation d'oxygène dissous

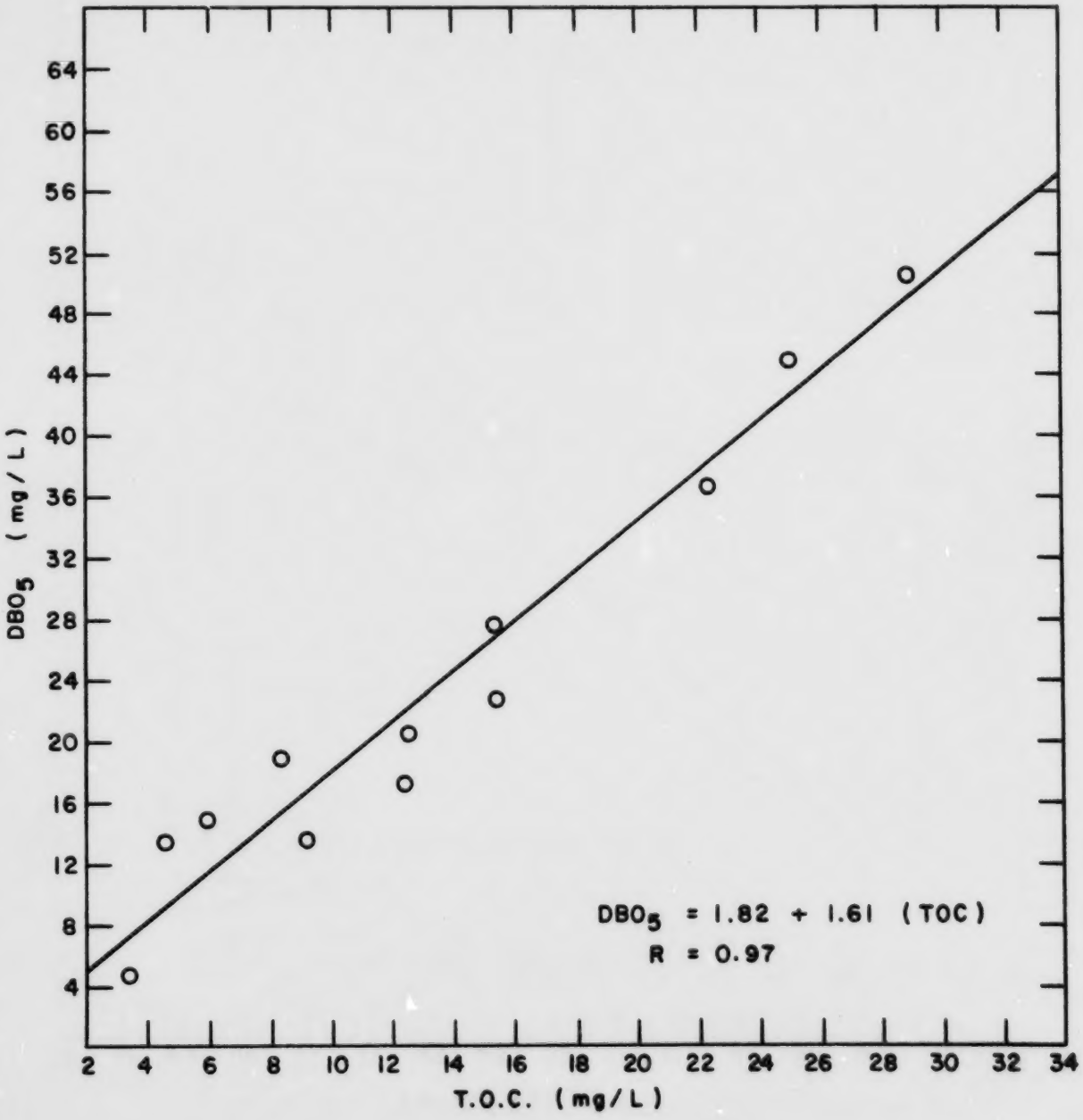

Figure 5. Relation TOC-DB05

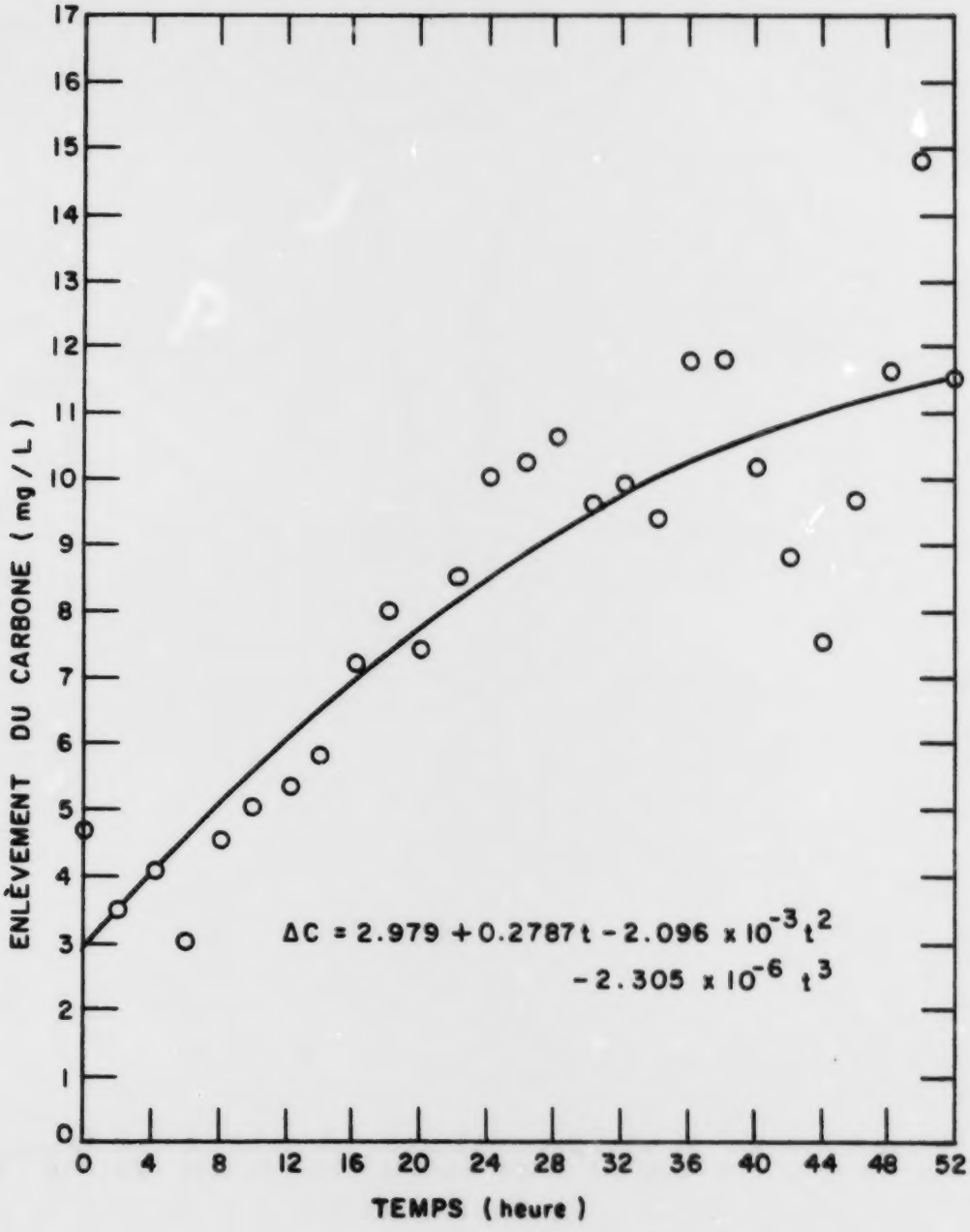

Figure 6. Deuxième essai: évolution de l'enlèvement du carbone

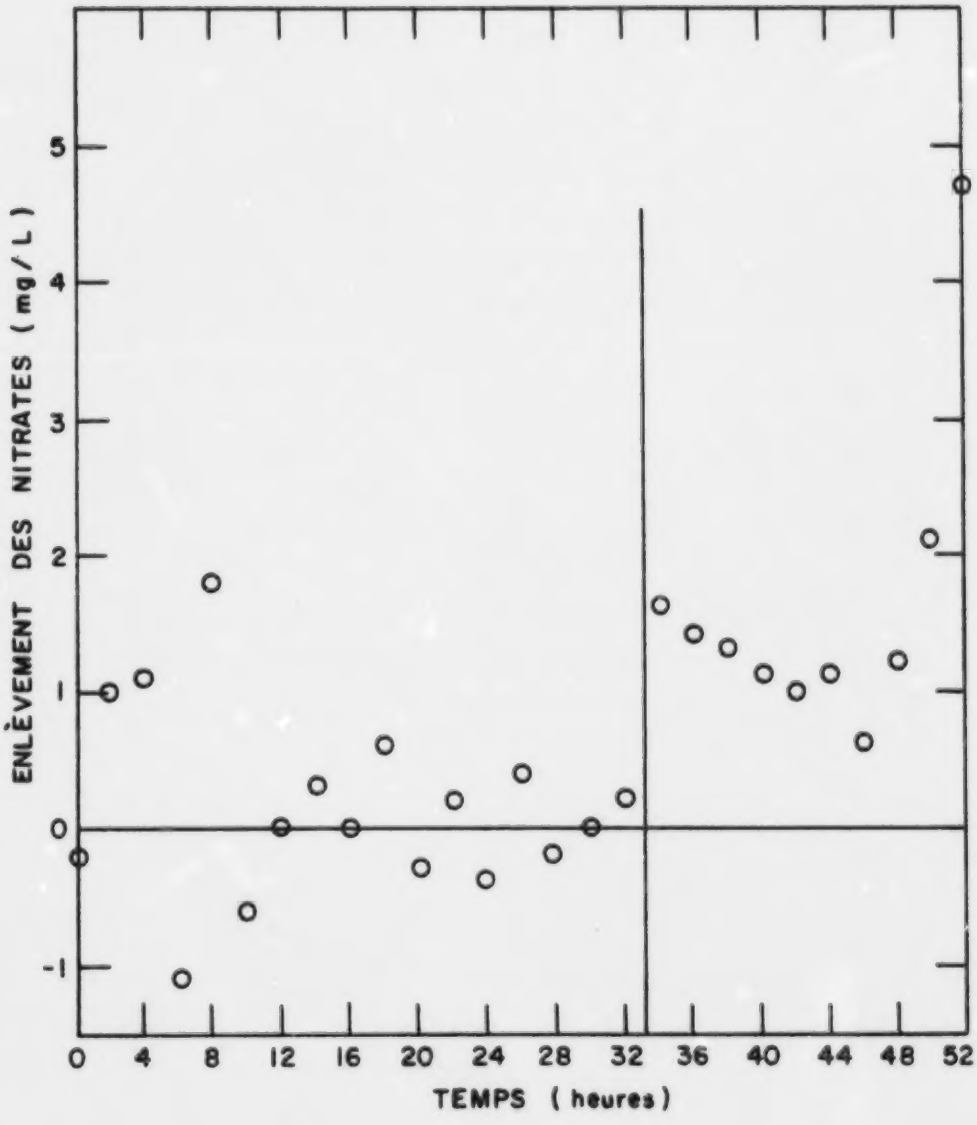

Figure 7. Deuxième essai: évolution de l'enlèvement des nitrates

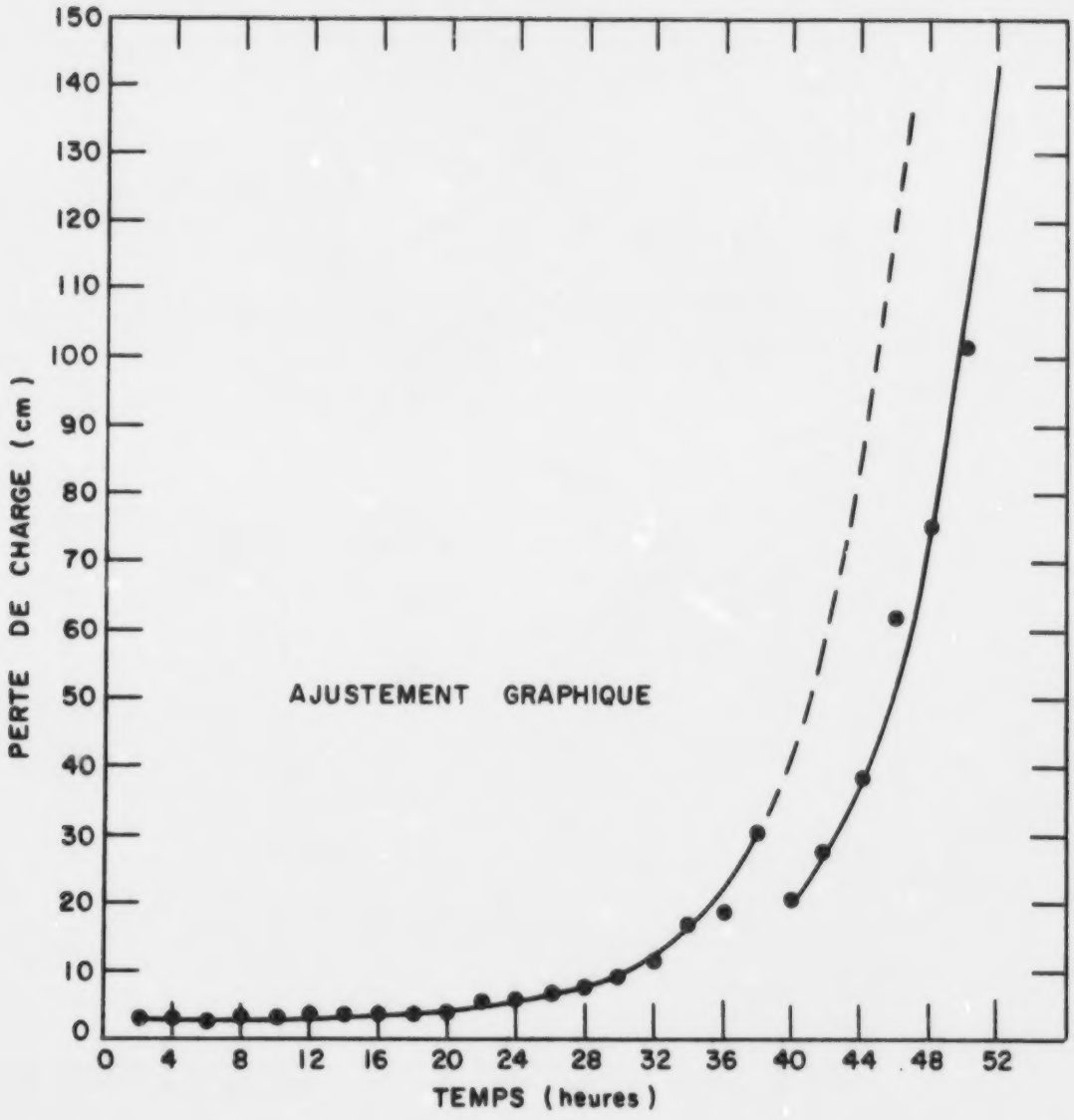

Figure 8. Deuxième essai: évolution de la perte de charge totale

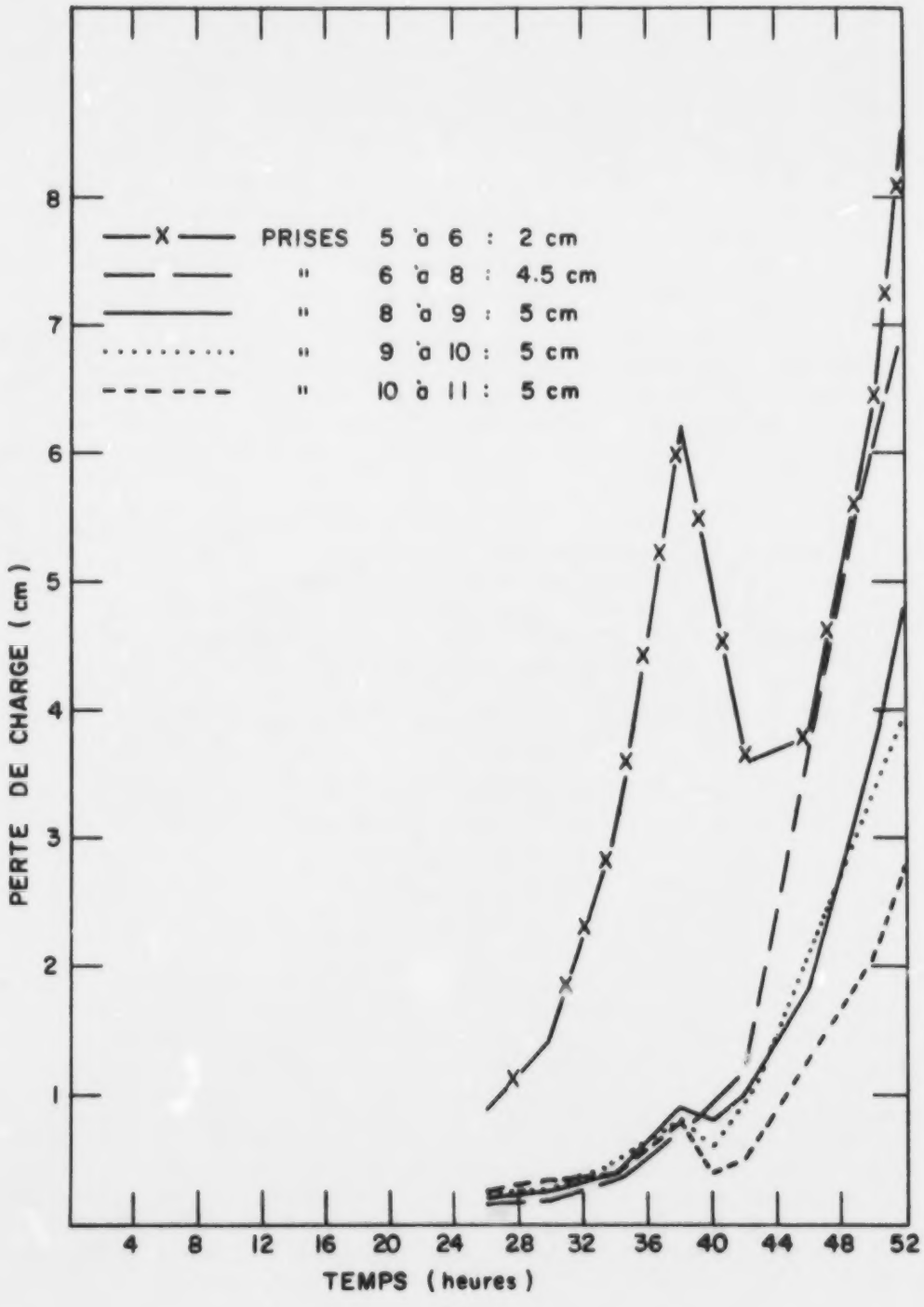

Figure 9. Deuxième essai: évolution de la perte de charge unitaire en différents points du filtre

pour l'enlèvement biologique de la DBO soluble, on a d'abord cherché à trouver les relations existant entre la biomasse présente dans le filtre et les différents paramètres de l'épuration: perte de charge, consommation d'oxygène, enlèvement du carbone, de la DBO et des nitrates.

Perte de charge vs biomasse

Le fait que la perte de charge continue d'augmenter de façon exponentielle alors que la biomasse plafonne, et même diminue, pose le problème de la relation entre les deux (2) phénomènes. L'augmentation de la perte de charge n'étant causée que par la croissance de la biomasse dans le filtre, comment expliquer que celle-là continue d'augmenter alors que celle-ci plafonne et même régresse.

Hoehn et Ray (5) ont montré que la densité d'un film biologique n'est pas constante. Dans la première phase de croissance du film, jusqu'à une épaisseur de 100 à 150 microns, la densité peut passer de 45 à 105 mg/ml. Aux environs de 150 microns, l'oxygène qui ne parvient plus aussi facilement aux couches inférieures du film devient le facteur limitant. Les microorganismes aérobies sont remplacés par des microorganismes facultatifs et anaérobies. La densité des microorganismes diminue et avec elle, celle du film biologique qui peut passer de 105 à 25 mg/ml. On assiste à un gonflement du film biologique dont le volume continue d'augmenter, et avec lui, la perte de charge, alors que son poids est stable ou même diminue.

C'est cette variation de la densité du film biologique qui explique l'allure particulière de la relation entre la biomasse et la perte de charge (figure 10). A un même poids de la biomasse peuvent correspondre plusieurs volumes et conséquemment plusieurs valeurs de perte de charge.

Une biomasse de 6,5 g/L peut causer une perte de charge de 9 ou de 100 cm dépendant de la phase de filtration dans laquelle on se trouve.

Consommation d'oxygène vs biomasse

Dans la corrélation entre la consommation d'oxygène et la biomasse, on peut délimiter trois (3) zones (figure 11). De 3 à 4 g/L de biomasse, il y a augmentation de la consommation d'oxygène alors que la biomasse est, à toutes fins pratiques, stable. La biomasse qui a résisté au lavage du filtre met un certain temps avant de commencer à se multiplier; c'est ce qu'on pourrait appeler le traumatisme du lavage.

Entre 4 et 7 g/L de biomasse, il y a une croissance parallèle de cette dernière et de la consommation d'oxygène dissous. Une corrélation linéaire rend bien compte du rapport consommation d'oxygène-biomasse dans cette zone.

Finalement, au-delà de 7 g/L de biomasse, c'est l'inverse du début, il y a accroissement de la biomasse alors que la consommation d'oxygène est stable. Ceci s'explique bien par le modèle proposé par Hoehn et Ray où, au-delà d'une certaine épaisseur du film, l'oxygène devient le facteur limitant. A partir de ce moment, la quantité de microorganismes aérobies se stabilise et l'épaisseur du film biologique s'accroit grâce à l'augmentation des organismes anaérobies et morts. Il y a donc croissance de la matière organique, mais plafonnement de la biomasse aérobie et de la consommation d'oxygène. Notons que l'équation polynominale ne rend pas très bien compte de la corrélation.

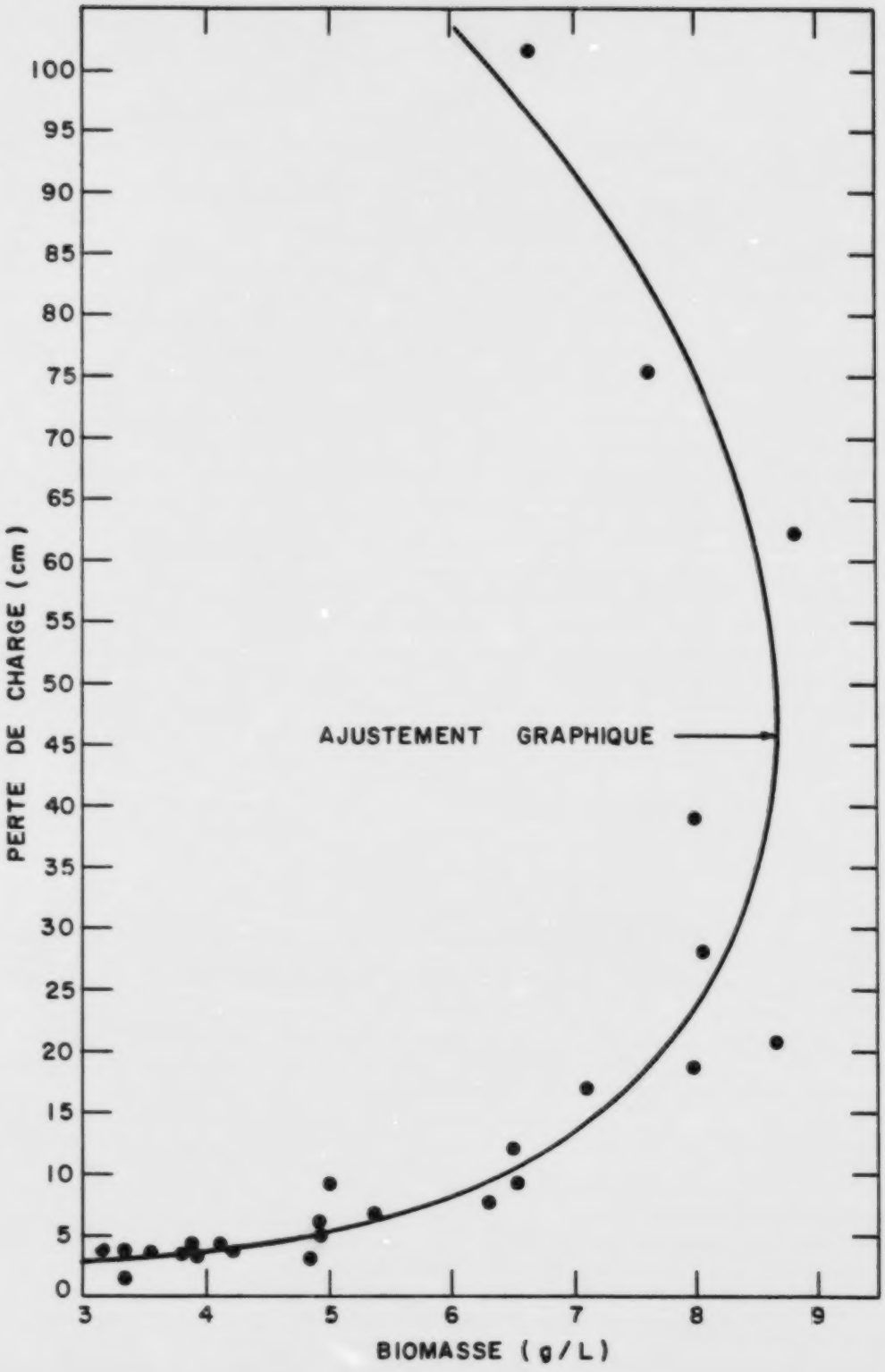

Figure 10. Deuxième essai: perte de charge vs biomasse (cycle complet)

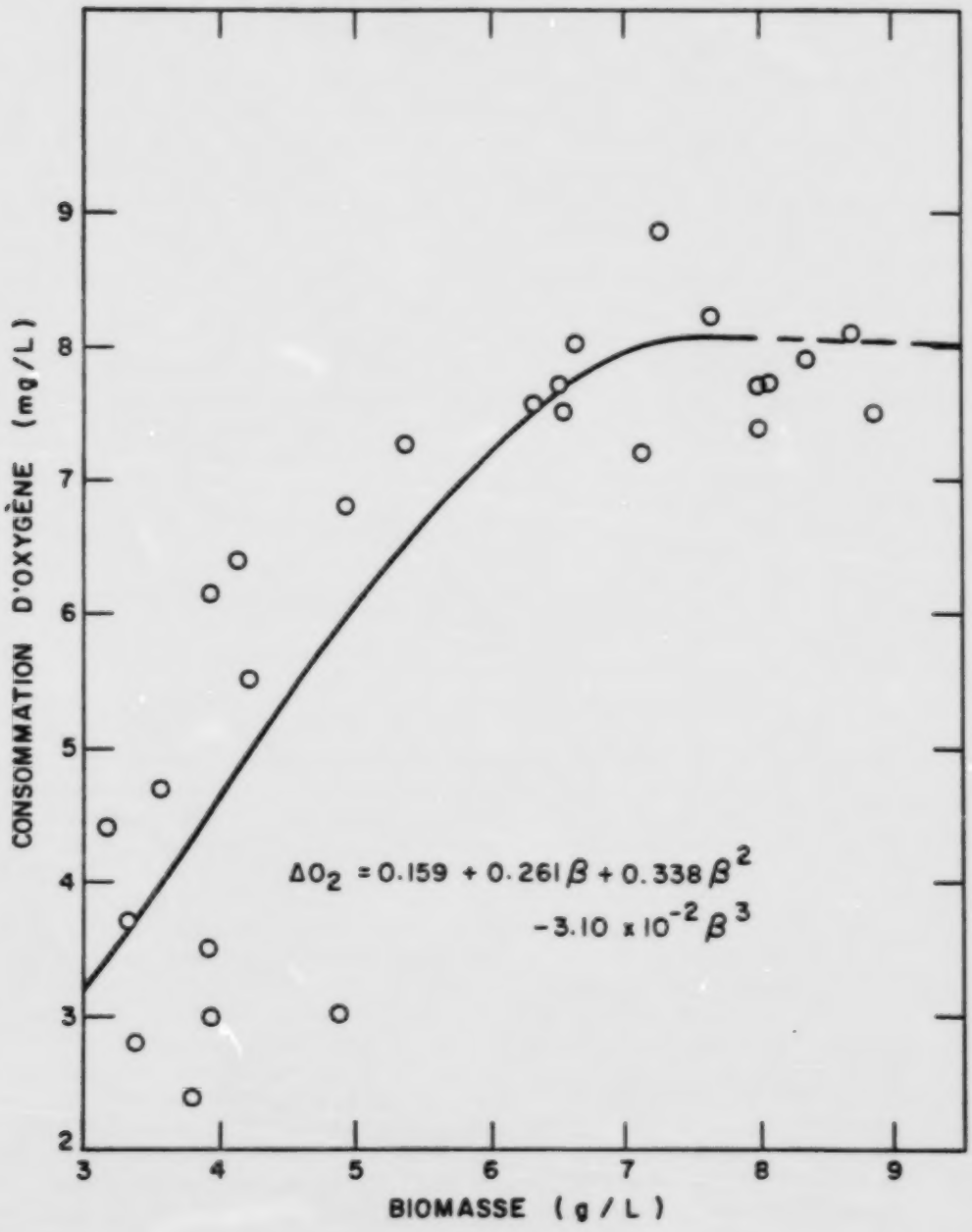

Figure 11. Deuxième essai: consommation d'oxygène dissous vs biomasse

Enlèvement du carbone vs biomasse

La relation entre l'épuration et la biomasse (figure 12) est similaire à celle entre la consommation d'oxygène dissous et la biomasse; ce qui va de soi, puisqu'il y a une forte corrélation linéaire entre la consommation d'oxygène dissous et l'enlèvement du carbone sauf dans les dernières heures du cycle (figure 13). Il faut noter ici qu'il y a de l'oxygène non dissous (sous forme de fines bulles) qui a été utilisé pour l'épuration. Ceci est évident si on considère que l'enlèvement du carbone organique a atteint 11 mg/L, ce qui correspond à 17,6 mg/L de DBO_5 (rapport de 1,6 entre les deux (2)) alors que la consommation correspondante d'oxygène dissous était de 8.

Si on considère que, par définition, il faut un peu plus de 1 mg/L d'oxygène pour l'enlèvement de 1 mg/L de DBO_5, on en conclue qu'il y avait une consommation d'oxygène non dissous supérieure à celle de l'oxygène dissous.

La faible corrélation entre la consommation d'oxygène dissous et l'enlèvement du carbone en fin de cycle peut s'expliquer, du moins en partie, par ce fait.

Enlèvement des nitrates vs biomasse

La corrélation entre l'enlèvement des nitrates et la biomasse (figure 14) est trop faible pour qu'on puisse en tirer une loi quelconque. On peut seulement dire qu'il ne semble pas y avoir d'enlèvement significatif des nitrates lorsqu'il y a moins de 7 g/L de biomasse dans le filtre, soit précisément au point où l'enlèvement du carbone plafonne. Ceci s'explique facilement du fait que la dénitrification est un processus qui s'effectue en absence d'oxygène. C'est donc au moment où la partie aérobie du film biologique atteint son épaisseur limite et qu'apparaissent à la partie inférieure les organismes facultatifs et anaérobies que l'enlèvement du carbone plafonne et que la dénitrification apparaît de façon évidente, bien que faible.

c) Performance globale de la filtration biologique

Au taux de filtration de 71,6 L/m^2-min., la filtration biologique par filtre à sable classique peut enlever de 40 à 50% de la DBO_5 soluble d'une eau usée contenant 20 à 25 mg/L de carbone organique ou 35 à 40 mg/L de DBO_5 soluble (tableau 3). Le taux d'enlèvement est fonction de la concentration à l'entrée, du taux de filtration et, comme on peut le voir par la comparaison des essais 1 et 2, d'un bon ensemencement biologique du filtre, d'un bon conditionnement aux conditions de la filtration. Le filtre était garni de 2500 à 8000 mg/L de biomasse, valeur comparable à celle des boues activées. L'enlèvement des nitrates est très faible (environ 10%) et ne se produit que dans la dernière partie du cycle, dans la partie la moins rentable à utiliser parce que la perte de charge augmente rapidement alors que l'épuration plafonne.

Bien que pour diverses raisons, le troisième essai n'a pas atteint la performance à laquelle on s'attendait, on peut raisonnablement penser que c'est dans cette gamme de 100 à 150 L/m^2-min. que se situe le taux optimal de filtration, celui qui permet l'utilisation la plus productive de la biomasse.

Les essais de filtration effectués ont montré que la façon la plus simple de suivre l'évolution de l'épuration est la mesure de l'oxygène dissous même si la corrélation n'est pas toujours parfaite entre les deux (2) paramètres.

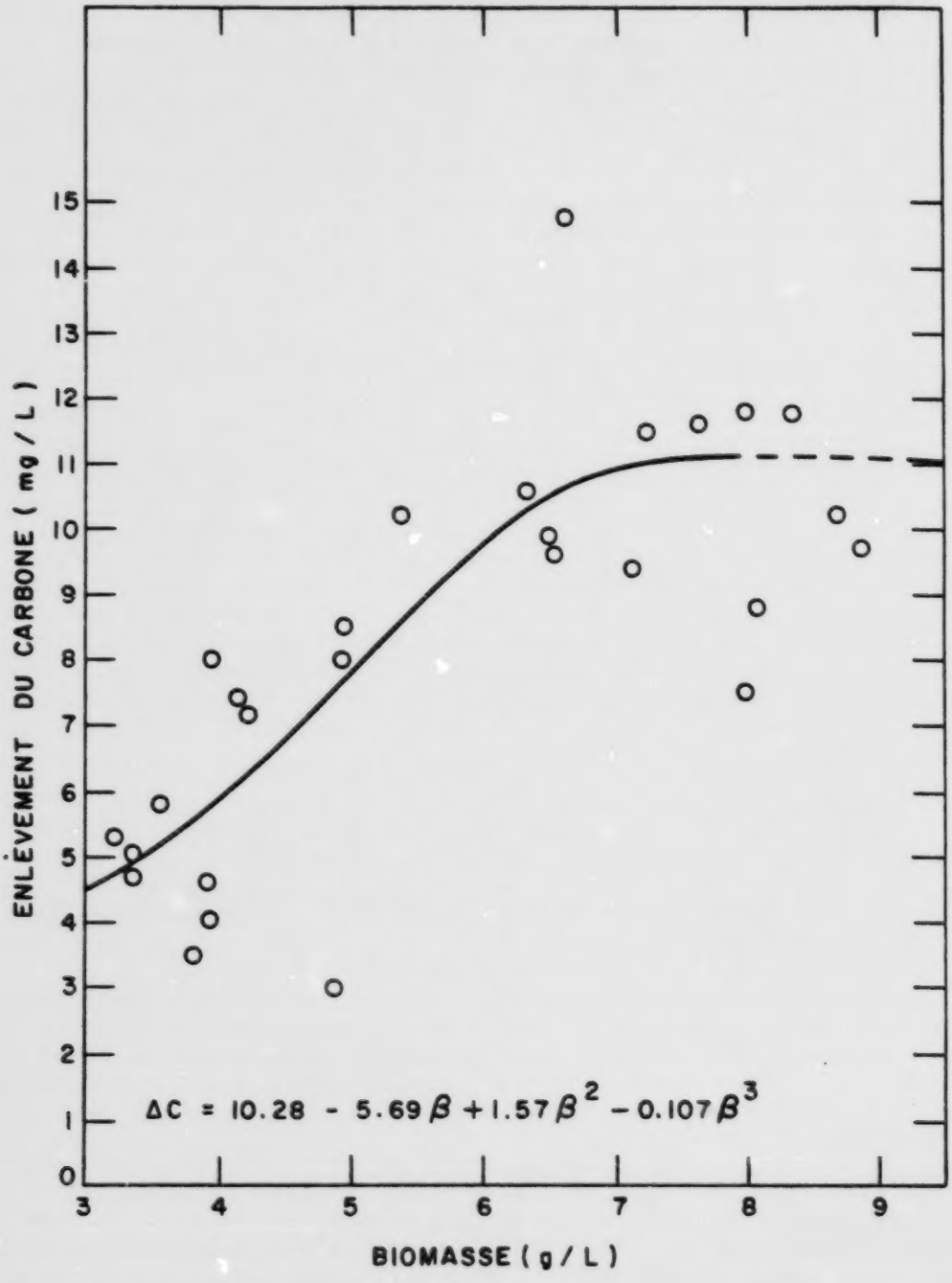

Figure 12. Deuxième essai: enlèvement du carbone vs biomasse

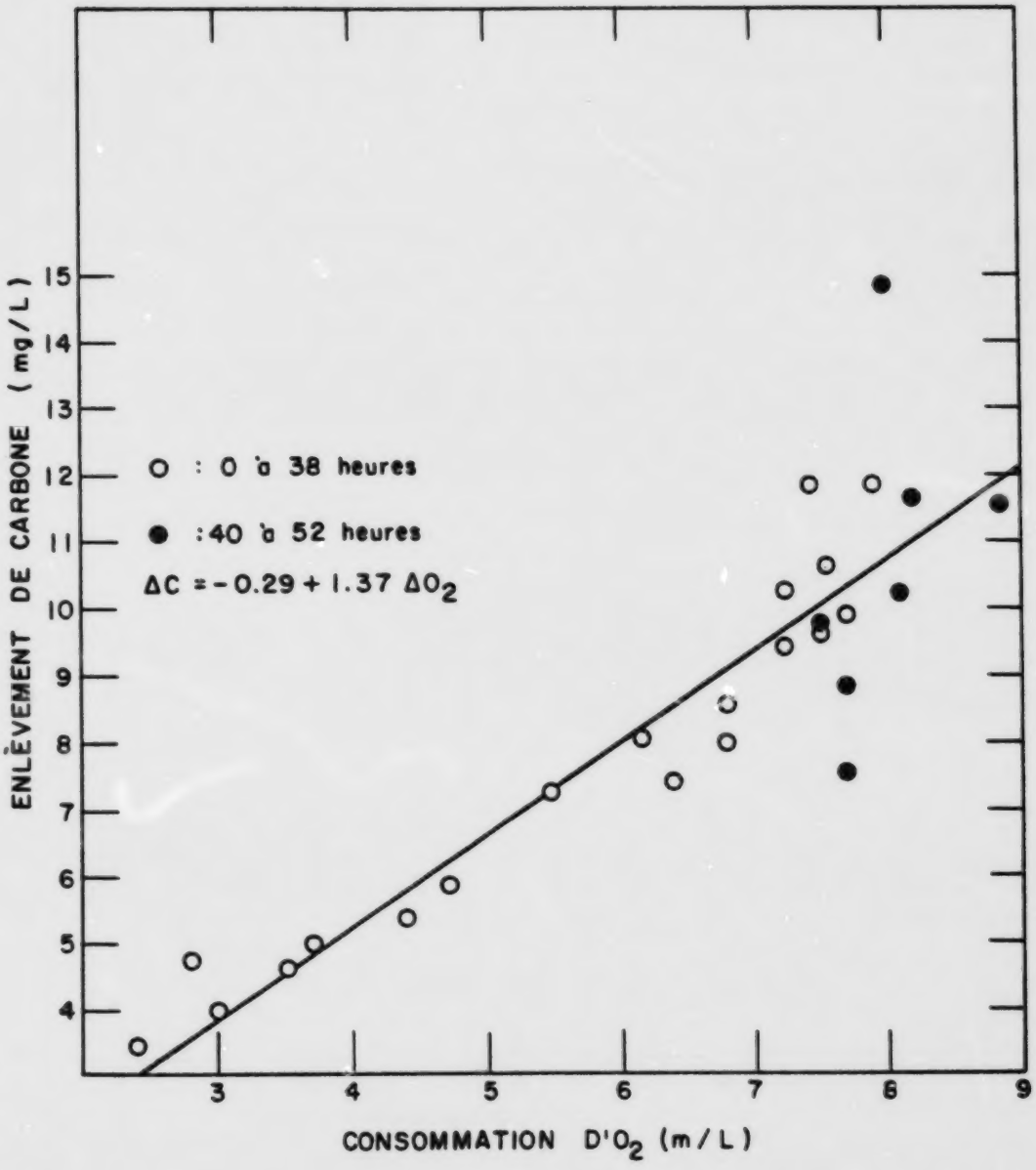

Figure 13. Deuxième essai: enlèvement du carbone vs consommation d'oxygène dissous

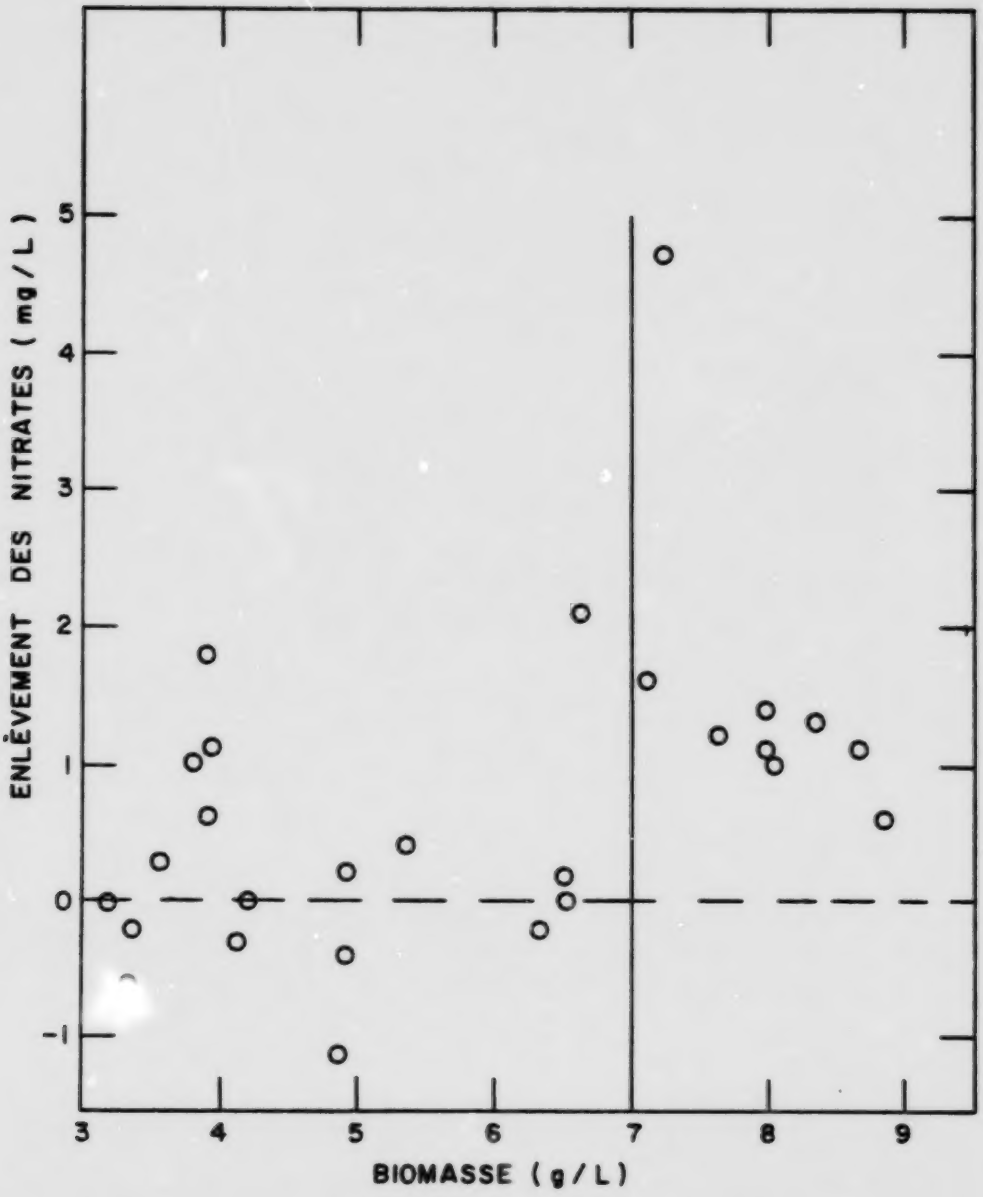

Figure 14. Deuxième essai: enlèvement des nitrates vs biomasse

TABLEAU 3. Valeur moyenne des différents paramètres pour les trois essais

Essais	Carbone				Nitrate			
	Entrée (mg/L)	Sortie (mg/L)	Enlèvement (mg/L)	Enlèvement (%)	Entrée (mg/L)	Sortie (mg/L)	Enlèvement (mg/L)	Enlèvement (%)
#1	23,2	12,0	11,2	48,3	9,0	7,8	1,2	13,3
#2	20,2	12,0	8,2	40,6	8,0	7,3	0,7	8,8
#3	16,6	10,1	6,5	39,2	7,2(1)	6,9(1)	0,3(1)	4,2(1)

(1) exclus la valeur de la 2ᵉ heure.

Essais	Oxygène dissous			Biomasse (g/L)
	Entrée (mg/L)	Sortie (mg/L)	Consommation (mg/L)	
#1	10,2	1,8	8,4	5,73
#2	9,9	3,7	6,2	5,74
#3	8,8	4,2	4,6	7,18

Lorsque les facilités existent, la mesure du carbone organique total s'avère une façon très commode de suivre l'épuration biologique à cause de l'étroite corrélation qui existe entre celui-ci et la DBO_5 pour un même substrat.

CONCLUSION

Les cycles de filtration se déroulent en trois (3) phases similaires aux trois (3) phases de la croissance bactérienne: une phase d'acclimatation ou de croissance lente, une phase de croissance rapide et une phase de stagnation ou de plafonnement.

La corrélation entre le poids de la biomasse et les autres paramètres de l'épuration (consommation d'oxygène, enlèvement de la DBO et perte de charge) n'est pas très bonne. Trois (3) raisons peuvent expliquer cet état de fait: la difficulté d'effectuer un échantillonnage représentatif, le fait que certains paramètres sont liés au volume plutôt qu'au poids et que la densité de la biomasse change au cours d'un cycle, et, finalement, le fait qu'il y ait, surtout à la fin du cycle, desquamation ou décrochage de biomasse.

On peut cependant dire que dans la phase centrale du cycle de filtration, la consommation d'oxygène et l'enlèvement de la DBO sont une fonction linéaire de la biomasse.

Quant à la perte de charge, le peu de corrélation qu'elle a avec le poids de la biomasse s'explique par le fait qu'elle est reliée au volume de la biomasse et que celle-ci change de densité au cours d'un cycle de filtration.

Les essais effectués ont montré que l'activité biologique dans un filtre à sable classique pouvait enlever de 40 à 50% de la DBO_5 soluble d'une eau usée qui en contient de 35 à 40 mg/L. Une telle performance remet en question l'attitude généralisée de considérer la croissance de matière biologique dans un filtre à sable comme une nuisance et ouvre la voie à de nouvelles recherches en vue de préciser et de compléter ces résultats.

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RATIONALISATION DES FACTEURS INFLUENCANT LA CONCEPTION DES OUVRAGES D'EPURATION PAR VOIE BIOLOGIQUE

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INTRODUCTION

La façon d'envisager la conception et l'opération des systèmes de traitement biologique a beaucoup évoluer du temps où on concevait les ouvrages principalement sur une base hydraulique. Maintenant on intègre des notions qui tiennent compte de la charge organique et de la cinétique gouvernant l'enlèvement du substrat et la croissance de la masse microbienne. On utilise une gamme de coefficients comme les coefficients de synthèse et de consommation d'oxygène.

Parfois des études en laboratoire ou pilote sont exécutées dans le but de développer des critères de conception. Dans tel cas, souvent les paramètres d'opération qui généreront ces valeurs de conception sont choisis de façon subjective et fréquemment au hasard. Quand on n'entreprend pas d'études mais qu'on doive concevoir quand même, les critères de conception sont arrêtés sur des considérations historiques ou après avoir fait quelques permutations et combinaisons de possibilités d'opération. On s'assure que le choix final des critères de conception ne dévie pas beaucoup des valeurs citées dans la littérature ou utilisées couramment. On utilise fréquemment dans ces calculs l'équation suivante (1) :

$$\frac{1}{\theta} = YU - k_d \qquad (1)$$

où θ = âge de boue, inventaire de biomasse dans le réacteur, divisé par la biomasse extraite du traitement quotidiennement, incluant la masse perdue dans l'effluent, d.

U = taux d'enlèvement du substrat par unité de biomasse, g/g-d.

$= Q (S_o - S)/XmV.$

Y = coefficient de synthèse, biomasse produite par masse de substrat assimilé, g/g.

k_d = coefficient de décomposition de la biomasse.

S_o,S = concentration de la matière organique totale dans l'affluent et soluble dans l'effluent, respectivement, mg/l.

Q = débit, l/d.

V = volume du réacteur, l.

Xm = Concentration de la biomasse active dans le réacteur, mg/l.

L'équation 1 suppose une droite dont le coefficient Y serait constant. En réalité on sait que ce coefficient n'est pas une constante; d'ailleurs Heukelekian et al (1951) parmis d'autres chercheurs ont démontré que le coefficient de synthèse variait inversement avec l'âge de boue. Vasicek (1982) concluait dans une étude sur l'opération d'un système de traitement par boue activée que les coefficients normalement supposés constants ne le sont pas. Il cite dans son étude des valeurs du coefficient de synthèse allant de 0.5 à 0.8 g de solides produits /g de DBO_5 enlevé. James (1982) faisant un tour d'horizon des méthodes de modelisation soulignait le danger de développer des modèles mathématiques qui ne reflètent pas la dynamique biologique. Il insistait sur la nécessité d'utiliser un coefficient de synthèse variable tout en tenant compte du rapport qui existe entre la croissance microbienne et l'utilisation du substrat.

Non seulement les coefficients de synthèse ne sont pas constants mais ils sont spécifiques à chaque eau usée, soit-elle de nature urbaine ou industrielle. La variabilité du coefficient de synthèse ne permet pas l'utilisation d'une droite tel qu'exprimé par l'équation en (1) car Y, la pente de cette droite, n'est pas constant. Cette constatation laisse un doute sur la manière de générer des chiffres de coefficients graphiquement.

Cette équation aussi ne donne pas d'indication sur la performance du système, car **U**, aussi connu comme le rapport **F/M**, ne tient pas compte de la qualité du floc et de la décantabilité de la biomasse qui en découlerait. En fait, on suppose un degré d'enlèvement et par calcul on détermine la masse biologique qui devrait résulter pour effectuer cet enlèvement. Il est aussi possible d'obtenir plusieurs combinaisons de **F** et de **M** pour le même rapport.

Finalement, la difficulté de prédire le coefficient de synthèse qui correspond à telle condition d'opération et à telle eau usée nous force à concevoir en choisissant des valeurs conservatrices ainsi que d'incorporer des marges de sécurité plus large que nécessaire.

Il est évident que ces questions ont des répercussions importantes sur les quantités de boues produites, le mode d'opération éventuel, la stabilité et le rendement, le dimensionnement et le degré d'optimisation de l'investissement dans les ouvrages d'assainissement.

L'objectif de cette recherche est de tenter d'évaluer l'interaction entre deux (2) principaux paramètres de conception et d'opération soit l'âge de boue et le temps de rétention hydraulique et de rationaliser leurs effets sur le rendement au moyen d'une surface de réponse. L'objectif secondaire est de démontrer que les données, qui peuvent sembler être disparates et difficiles à expliquer, prennent un sens nouveau lorsque visualisées sur une surface.

APPROCHE

Le concept de surface de réponse repose sur le fait que, pour chaque combinaison d'âge de boue et de temps de rétention hydraulique pour une eau usée d'une concentration en matières organiques donnée, on obtient une réponse spécifique du système biologique. Tous les autres facteurs importants à la croissance biologique sont maintenus constants. Ce concept a été développé en examinant les données recueillies de diverses études sur le traitement d'effluents urbain et industriel.

Reconnaissant l'importance d'assurer aux microorganismes un milieu dans lequel des conditions favorables de croissance existaient, comme il est bien recommandé dans la littérature (1,5), toutes ces études étaient menées sous des régimes de mélange complet, de suffisance d'éléments nutritifs et d'oxygénation, de pH adéquat et de température contrôlée. Le régime hydraulique pouvait être caractérisé de constant en autant que les débits étaient maintenus constants; les concentrations des affluents bruts n'étaient pas ajustées et, par conséquent, des affluents de charges variables étaient traités. Les détails de ces études ont déjà été publiés (6,7 et 8).

DEVELOPPEMENT DE LA SURFACE

Lorsqu'on examine de près l'équation 1, on note que les deux (2) paramètres d'importance contrôlables, qui peuvent être utiles à la conception et à l'opération, sont le temps de rétention hydraulique (V/Q) et l'âge de boue. Ces paramètres spécifient les dimensions des ouvrages. Le rapport F/M, défini comme matière organique enlevée par la masse biologique, peut être décomposé pour devenir:

$$F/M = \frac{(S_o - S)(Q)}{(Xm)(V)} \qquad (2)$$

où Q = débit moyen, l/d.

V = volume du bassin d'aération, l.

S_o = concentration de la matière organique dans l'affluent, mg/l.

S = concentration de la matière organique dans l'effluent, mg/l.

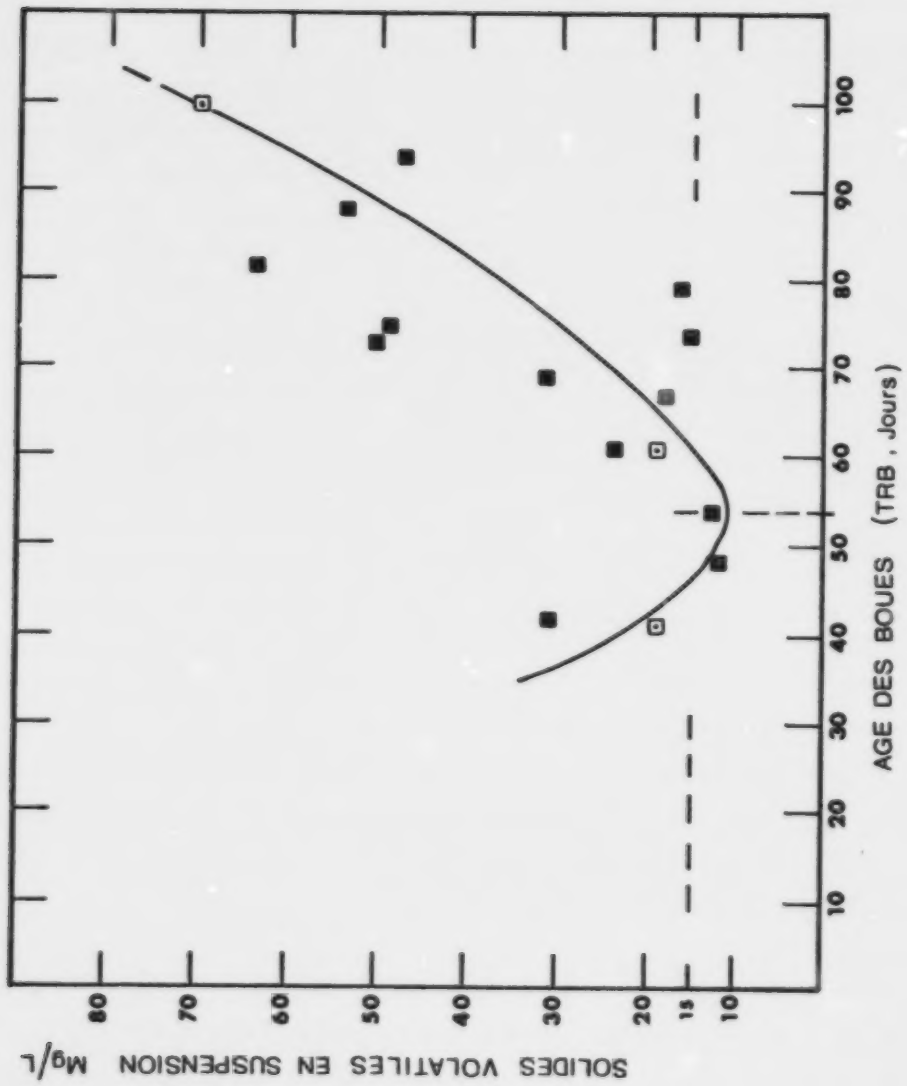
on retrouve $t = V/Q$
= temps de rétention hydraulique, d.

Vu que le rapport F/M en soi peut difficilement être un paramètre de conception, mais surtout un guide, on retombe nécessairement sur le temps de rétention hydraulique qui fixe le volume du bassin d'aération.

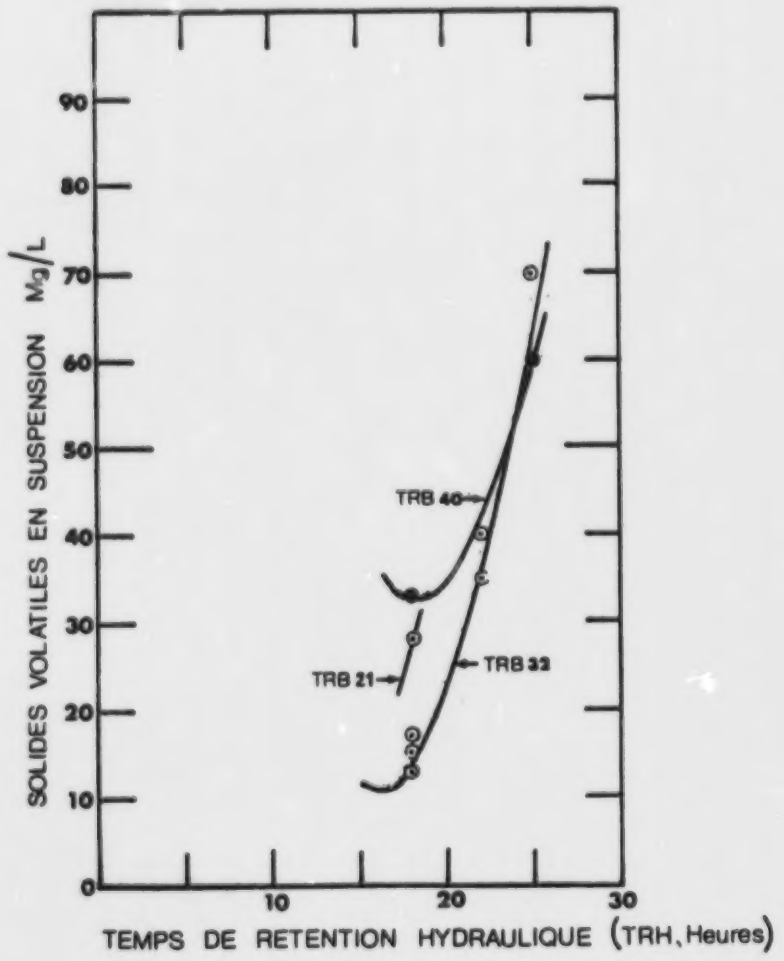
L'autre composante de l'équation 1 est l'âge de boue qui peut être contrôlé aisément lors de l'opération et qui définit aussi les quantités de boues à extraire.

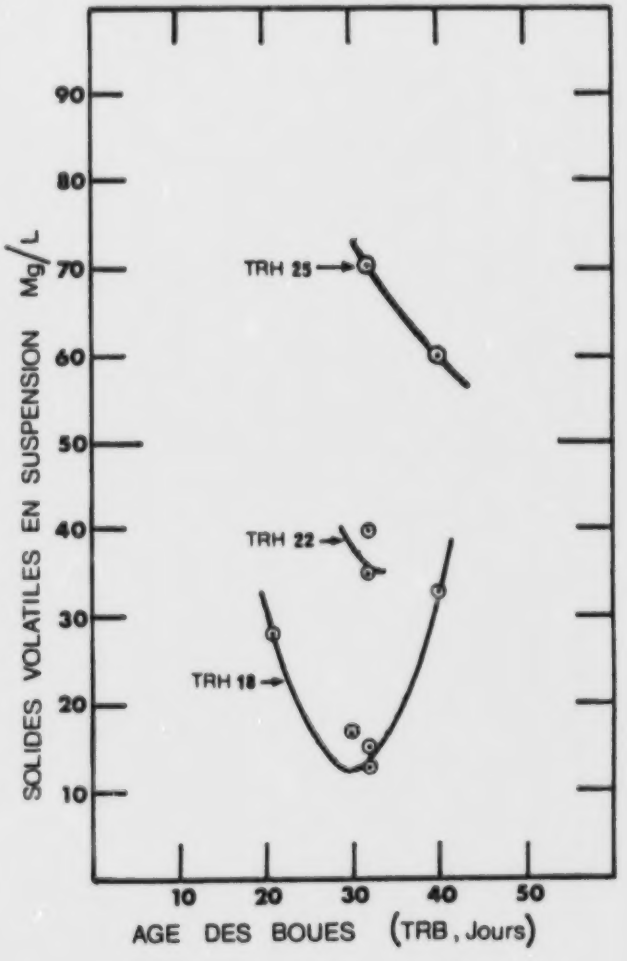
L'âge de boue par définition fixe le temps de séjour d'une biomasse dans un système biologique et, indirectement, la qualité de cette boue. La littérature encore nous indique qu'une boue trop jeune a des propriétés de flocculation relativement mauvaises, exemple: le bulking, et les systèmes opérant à des âges de boues trop élevés, sont caractérisés par des propriétés de défloculation produisant un floc dispersé.

L'inter-relation entre le temps de rétention hydraulique, l'âge de boue et la réponse du système est très évidente lorsqu'on examine le graphique 1. Ce graphique d'un effluent de fromagerie (6), montre que variant l'âge de boue, en maintenant le temps de



GRAPHIQUE 1 — Effet de l'âge des boues sur le rendement d'un traitement.





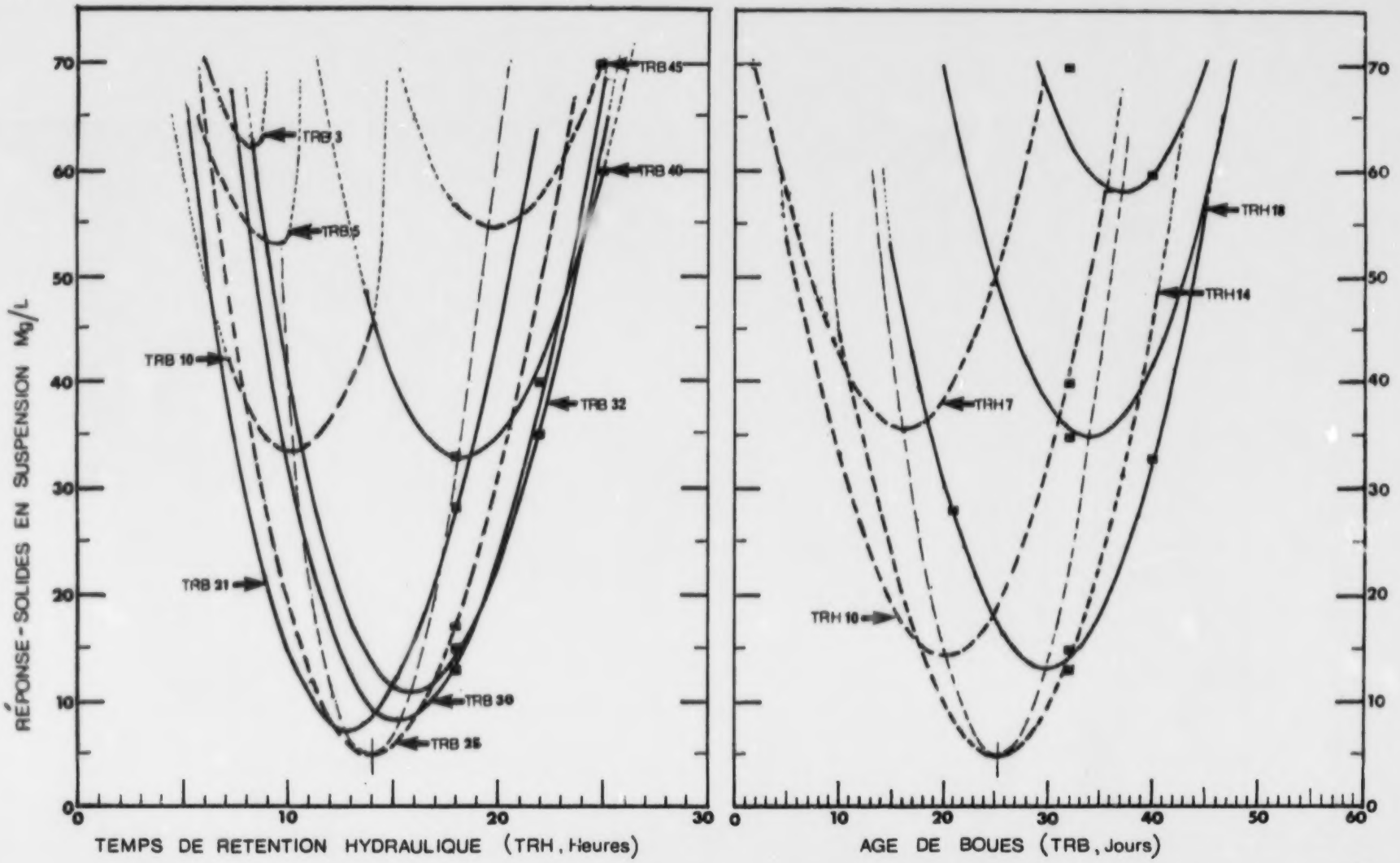
GRAPHIQUE 2 — Réaction d'un système biologique aux variations des paramètres de contrôle.

rétention hydraulique fixe, la concentration de l'effluent en solides volatiles en suspension passe par un minimum en deça duquel la qualité de l'effluent se détériore graduellement. On fait la même observation après l'examen de la réponse d'un système traitant un effluent d'usine de textile où l'âge de boue était maintenu fixe et le temps de rétention hydraulique était varié (8). Le graphique 2 montre cet exemple où l'influence de chacun des deux (2) paramètres est claire. Le graphique 2 montre aussi que les autres conditions expérimentales étudiées ont donné des résultats qui ne se trouvent pas sur la même courbe indiquant ainsi qu'ils appartiendraient à d'autres courbes. L'observation de ces graphiques nous permet de faire quelques constatations importantes:

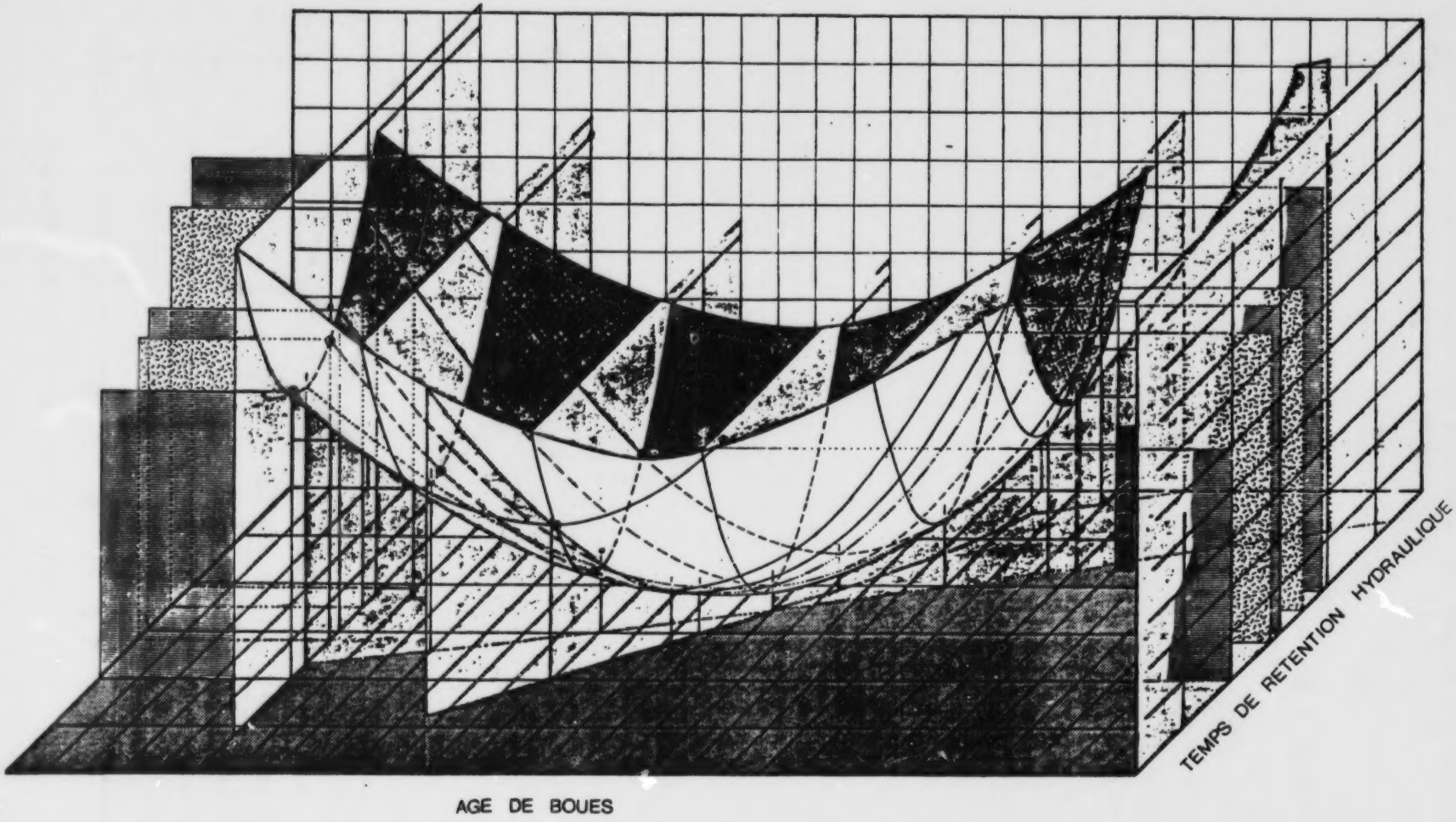
1. Si on maintenait l'âge de boue ou le temps de rétention hydraulique fixe en faisant varier l'autre paramètre sur une gamme de valeur, on obtiendrait des familles de courbes semblables à celles présentées aux graphiques 1 et 2.

2. Ces courbes spécifient les conditions d'opération qui donneraient un rendement attendu. Par exemple, au graphique 2b, afin d'obtenir un effluent ayant 20 mg/l de solides volatiles en suspension à un temps de rétention hydraulique de 18 heures, il faudrait opérer soit à 24 ou 36 jours d'âge de boue. Cette performance ne serait pas obtenue à d'autres âges de boue avec ce régime hydraulique.

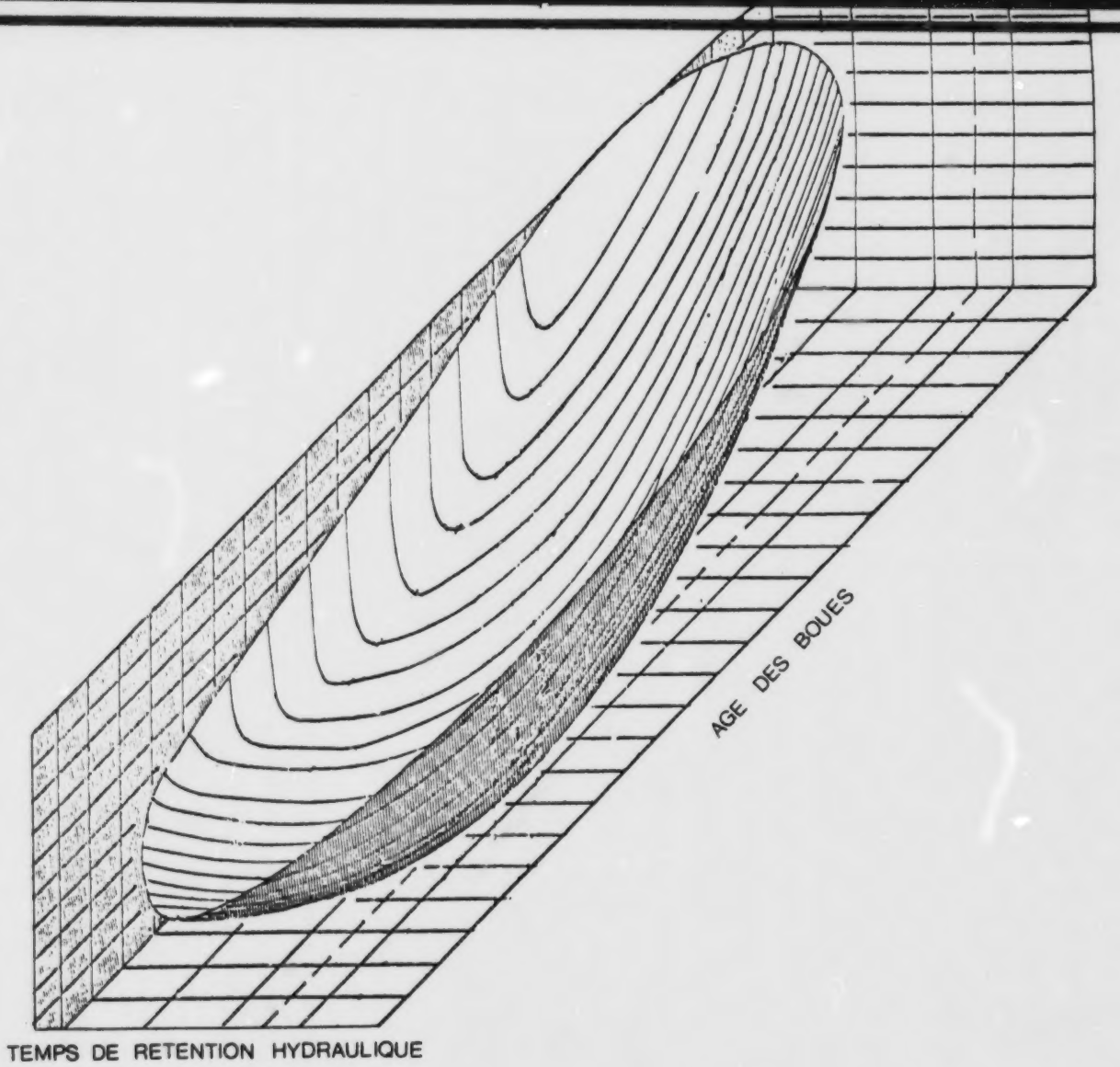
3. Si un nombre suffisant de données étaient obtenus à différents âges de boues et temps de rétention hydraulique, on obtiendrait des familles de courbes semblables à celles des graphiques 1 et 2 qui caractériseraient le comportement du système biologique. La figure 3 montre l'allure qu'auraient



GRAPHIQUE 3 – Représentation des courbes de réponse en fonction des paramètres de contrôle.



GRAPHIQUE 4 — Représentation de la construction d'une surface de réponse.



GRAPHIQUE 5 – Illustration de l'allure d'une surface de réponse.

ces courbes. Les deux (2) graphiques de la figure 3 proviennent de la projection d'une représentation dans l'espace de la surface de réponse. Les figures 4 et 5 montrent une illustration de l'allure qu'aurait une telle surface.

CARACTÉRISTIQUES DE LA SURFACE DE RÉPONSE

Étant donné que les représentations sur les plans **XZ** et **YZ** de la figure 3 sont de formes paraboliques, la surface elle-même doit avoir cette forme générale. Chacune de ces courbes a un minimum local et la surface elle-même aurait un minimum global où le rendement serait optimal. À partir de cette surface on peut faire quelques observations intéressantes sur des tendances importantes:

1. Une orientation vers un temps de rétention hydraulique et un âge de boue court produirait un effluent élevé en matières en suspension, en DBO_5 ou DCO (fraction soluble et totale). La performance du système deviendrait de plus en plus mauvaise et ultimement il n'y aurait plus de traitement.

2. Une orientation à l'autre extrême vers un temps de rétention hydraulique élevé et un âge de boue élevé tendrait à produire un effluent chargé en solides en suspension (pin floc) et en DCO OU DBO_5 totales. La fraction soluble de la DCO ou de la DBO_5 aura déjà été éliminée.

3. À l'intérieur de ces deux (2) extrêmes d'âge de boues et de temps de rétention hydraulique on obtient une performance intermédiaire jusqu'à l'atteinte de l'optimum.

4. Cette surface définie une enveloppe à l'extérieur de laquelle il est impossible d'opérer le système.

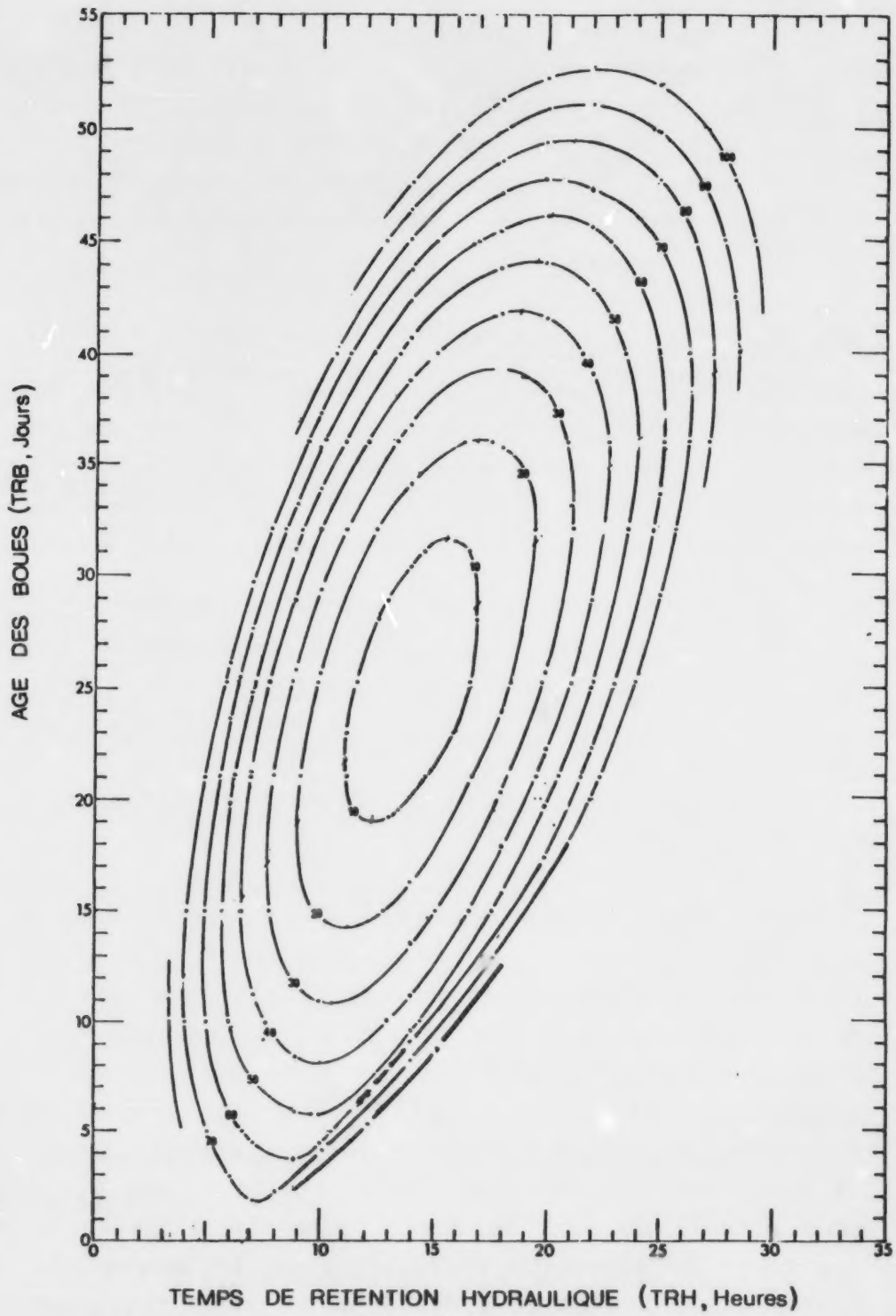
5. Dépendamment de la courbature de la surface on aurait des zones d'instabilités opérationnelles (courbature prononcée et qualité d'effluent difficile à maintenir) ou des zones d'insensibilités (surface relativement plate, où des changements de conditions d'opération n'affectent pas sensiblement le rendement).

6. Il existe une combinaison unique d'âge de boue et de temps de rétention hydraulique pour chaque valeur de réponse sur cette surface.

7. Il y aurait une surface spécifique à chaque eau usée à traiter.

8. Si la concentration de l'eau brute changait à l'équilibre, la liqueur mixte s'ajusterait à une valeur correspondante et on obtiendrait la même réponse pourvu que le temps de rétention hydraulique et l'âge de boue ne soient pas modifiés (tableau 2). Cette surface, donc, définirait les possibilités de conception et d'opération et éliminerait les choix impossibles.

Lorsque cette surface est tranchée à des niveaux de réponses choisis, on obtient une gamme de courbes et d'ellipses d'iso-réponses. Chacune de ces courbes et ellipses spécifie un nombre de combinaisons de temps de rétention hydraulique et d'âge de boue pour lesquels on aurait la même performance. Par conséquent, il devient possible d'effectuer une évaluation économique de plusieurs combinaisons tout en étant assuré d'avoir une qualité d'effluent connue, comme le montre la figure 6. Par exemple, pour obtenir 20 mg/l il serait possible d'opérer en maintenant des temps de rétention hydraulique de 10 heures, 12 heures, 15 heures et 18 heures et à des âges de boues correspondants de 15 jours, 30 jours, 35 jours et 24 jours (Il faut noter que les



GRAPHIQUE 6 – Courbes d'iso-réponses en fonction des deux paramètres de contrôle

chiffres montrés sur le graphique 6 sont extrapolés d'une étude de traitabilité (8) pour fin d'illustration).

Cet outil permet de manipuler l'opération pour rencontrer les changements de régime; il permet aussi de choisir des conditions d'opération sur une base rationelle et à la limite, il permet de déterminer le besoin de recommander une expansion à la capacité du système.

Incidemment, Cashion et Keinath (1983), qui évaluaient l'effet de trois (3) variables (temps de rétention hydraulique, âge de boue et taux de débordement) sur l'efficacité de la clarification dans le traitement par boue activée, ont défini une surface de réponse qui aurait la forme d'une selle de cheval. Leur modèle mathématique indiquait qu'une hausse ou une baisse simultanée de l'âge de boue et du temps de rétention hydraulique résultait en une augmentation des solides en suspension dans l'effluent. Ces résultats concordent avec les conclusions présentées plus tôt. Leur surface en forme de selle mène à des conclusions différentes que celles présentées ici quant à l'augmentation d'une variable et à la baisse de l'autre, et vice versa. Ils concluaient qu'une amélioration de la performance en résultait. Dans la présente étude, on arrive à la conclusion que des performances intermédiaires sont obtenues jusqu'à l'atteinte d'une combinaison optimale. Ces auteurs ont aussi conclu que le taux de débordement exerçait une faible influence sur la concentration des solides en suspension dans l'effluent pour des taux inférieurs à 40 m^3/j. Dans la présente étude, les décanteurs utilisés étaient surdimensionnés de sorte que les taux de débordement étaient effectivement sans effet. La qualité de l'effluent reflétait essentiellement l'effet de la qualité du floc sur la performance.

RELATION ENTRE LA SURFACE DE RÉPONSE ET LE COEFFICIENT DE SYNTHÈSE

Le coefficient de synthèse qui découle de cette surface est un coefficient net car dans la mesure de quantité de boue produite par quantité de matières organiques enlevées, les effets de prédation et de respiration endogène sont inclus.

La surface de réponse prévoit qu'il existe une réponse pour chaque combinaison d'âge de boue et de temps de rétention hydraulique. Le coefficient de synthèse net, qui est la mesure de la quantité de boue produite (boue extraite + boue perdue à l'effluent) par quantité de substrat enlevé (DCO_t - DCO soluble dans l'effluent), est directement influencé par les choix d'âge de boue et de temps de rétention hydraulique. L'équation 1 ou sa variation en fonction d'un coefficient net (5)

$$t = Y_{net} \frac{\Delta DCO}{X} \theta \qquad (3)$$

ne permet pas la prévision des valeurs (10).

Par conséquent, sans expérimentation, il est difficile de prévoir la performance d'un système pour un choix de t et de θ, et en même temps de savoir à quelle valeur d'équilibre s'établira la liqueur mixte X. Le choix d'une valeur de Y net arbitraire comme il est couramment fait et le choix de t et de θ auxquels ne correspond pas Y net introduit des erreurs importantes. Par exemple, une fois une station d'épuration construite, le t est fixé. Afin d'obtenir le meilleur rendement possible dans cette condition, l'opérateur devra manipuler l'âge de boue. Il pourrait se retrouver dans une situation où sa marge de manoeuvre est très étroite, ne lui permettant pas de corriger son état de fait. Le coefficient de synthèse utilisé lors de la conception et sur lequel des équipements et des dimensionnements auront été faits ne correspondrait pas nécessairement à celui mesuré sur le

terrain. Celui-ci aurait des coordonnées dont la réponse correspondrait à une valeur différente de celle anticipée dans la conception originale. Aussi, si l'on examine de près la surface de réponse, on note que le rendement pourrait se situer bien à l'écart du rendement optimal économique défini par les courbes d'iso-réponse (figure 6). Ainsi, des ouvrages sur-dimensionnés ou sous-dimensionnés pourraient en résulter et, advenant des changements de conditions tel que le débit ou la charge organique, on serait dans l'obligation de faire des modifications importantes et coûteuses qui aurait pu être évitées. Ces modifications pourraient être soit l'addition de structures nouvelles, soit l'abandon de structures existantes. Vasicek (1982) en est arrivé à cette conclusion au bout de trois (3) années d'évaluation d'une station d'épuration.

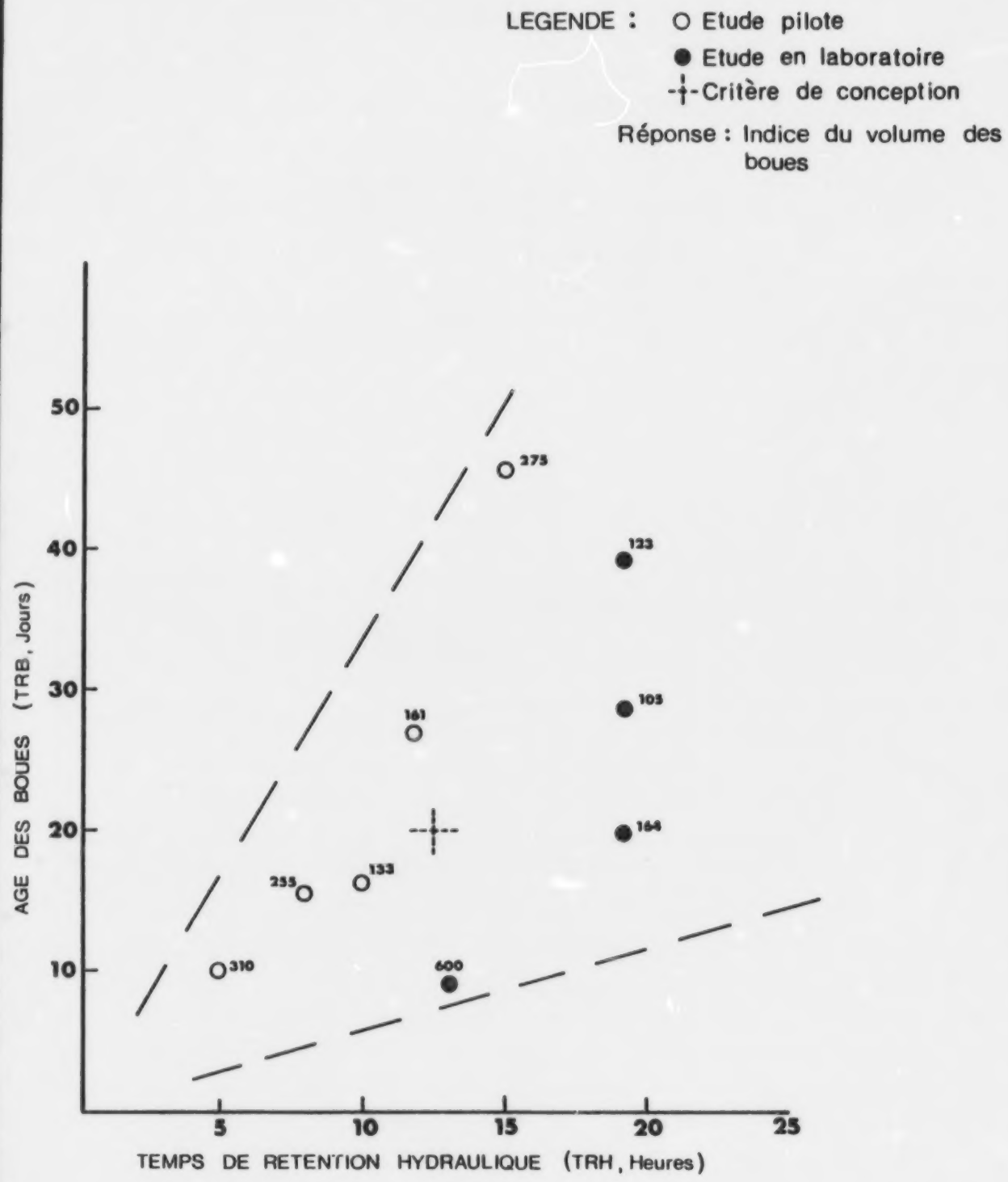
Ce concept confirme que le coefficient est variable et qu'à chaque condition d'opération un coefficient particulier découlerait. Si l'on examine à nouveau l'équation 3, on trouve que le rapport t/θ peut être maintenu constant en changeant t proportionnellement à θ. Ainsi, pour une gamme de valeur t et θ, le produit $Y_{net} \frac{\Delta DCO}{X}$ doit être constant. Il est clair que Y_{net} varie inversement selon les variations de Δ DCO et de X. Le produit $Y_{net} \frac{\Delta DCO}{X}$ est aussi la pente de la relation 3 qui donne une droite passant par l'origine. Vu que cette pente peut être manipulée à volonté, on obtient une gamme de droite formant un faisceau de conditions d'opération auxquelles correspondraient des valeurs de Y_{net} et de réponses propre à chaque mode d'opération. Comme on l'a dit plus tôt, il est impossible d'opérer un système avec recirculation de boue à des valeurs approchant des conditions de lessivage et espérer de faire varier l'âge de boue en maintenant le temps de rétention constant. Une fois les conditions de lessivage atteintes, soit des âges de boues et temps de rétention hydraulique très faibles, on se trouve à opérer un système sans recirculation de boue où $t = \theta$ et $Y_{net} = X/\Delta DCO$.

VÉRIFICATION DU CONCEPT DE LA SURFACE DE RÉPONSE - DEUX CAS TYPES

1. Étude de traitabilité des eaux usées de Granby (1979 à 1980)

Cette étude a été menée en deux (2) phases: la première phase était exécutée en laboratoire et l'alimentation provenait d'un échantillon composé prélevé trois (3) fois par semaine sur les principaux émissaires de la ville. La deuxième phase était exécutée sur le terrain en usine pilote et l'alimentation était prise directement dans l'égout à la chambre de déversement de l'intercepteur et du collecteur St-Urbain. Les eaux usées non comprises dans ces égouts étaient manuellement ajoutées. Dans les deux (2) cas, les eaux usées étaient décantées avant d'être alimentées aux bioréacteurs. Le tableau 1 résume les principaux résultats pertinents à la présente étude. La similitude du rapport DCO soluble /DBO_5 soluble indique bien que l'eau usée n'est pas différente dans les deux (2) phases de l'étude. Malheureusement, les DCO totales correspondant aux valeurs de DCO solubles de l'étude en laboratoire ne sont pas disponibles.

Malgré cela, il est permis de croire que les caractéristiques de ces deux (2) effluents sont sensiblement les mêmes. Le tableau 1 indique qu'à première vue il n'y a aucun rapport entre ces deux (2) phases de l'étude mais si on trace sur le même graphique les conditions d'opération et l'indice de volume de boue comme réponse des deux (2) essais, on observe une cohésion remarquable des résultats en fonction d'une surface de réponse. La figure 7 montre précisément cette constatation. On remarque que les valeurs des indices de boues élevées (reflètant une décantation détériorée) se situent vers l'extérieur de faisceau défini plus tôt. C'est le cas des valeurs de 310, 255, 275 et 600.



GRAPHIQUE 7 – Evolution de l'indice du volume des boues en fonction d'une surface de réponse

Tableau 1

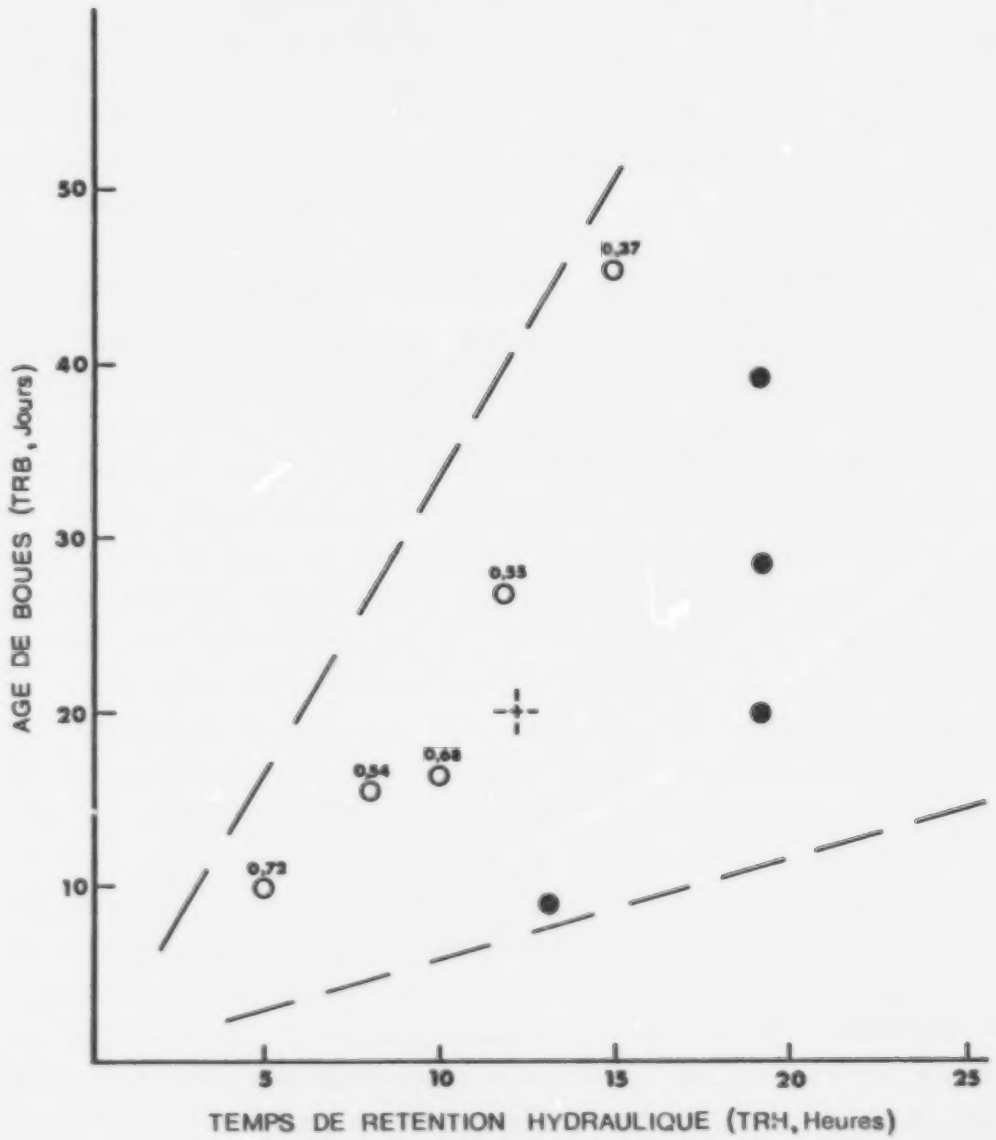
a) Résultats de l'étude en laboratoire des eaux usées de Granby

Âge de boues, d	9,2	19,8	28,9	39,1
Temps de rétention hydraulique, h	13,1	18,8	18,7	18,4
I.V.B. ml/g	600	164	105	123
DBO_5 soluble, mg/l (affluent)	82	76	76	76
DCO soluble, mg/l (affluent)	193	178	178	178
DCO soluble, mg/l (effluent)	42	25	20	23
Rapport DCO soluble/DBO_5 soluble dans l'affluent	2,35	2,34	2,34	2,34

b) Résultats de l'étude pilote des eaux usées de Granby

Âge de boues, d	45,5	27	16,4	15,6	9,8
Temps de rétention hydraulique, h	15	12	10	8	5
I.V.B. ml/g	275	161	133	255	310
DCO_t affluent/effluent, mg/l	198/34	180/34	169/31	205/31	162/37
DCO soluble, affluent/effluent, mg/l	114/22	150/26	108/25	123/24	94/28
DBO_5 totale affluent/effluent, mg/l	87/3	86/4	79/3	85/3	57/4
DBO_5 soluble, affluent/effluent, mg/l	57/2	59/2	48/2	54/2	29/2
Rapport DCO soluble/DBO_5 soluble (affluent)	2	2,54	2,25	2,28	3,2
Y_{net} base DBO_5 totale - DBO_5 soluble	0,37	0,55	0,68	0,54	0,72
Y_{net} base DCO totale - DCO soluble	0,17	0,30	0,36	0,24	0,30

LEGENDE : O Etude pilote
● Etude en laboratoire
-┼-Critère de conception
Ynet f(DBO)



GRAPHIQUE 8 – Représentation du coefficient net de synthèse en fonction d'une surface de réponse

Plus on se rapproche vers le creux de la surface, plus l'indice s'améliore indiquant ainsi une tendance vers un optimum. Les valeurs dérivées des tests en laboratoire concordent avec les valeurs obtenues des essais pilotes. Aussi, conformément à la notion des coupes d'iso-réponses, on peut voir que si un nombre plus grand de points expérimentaux étaient pris, il aurait été possible de tracer les coupes de la figure 6.

Le tableau 1 montre aussi les valeurs des coefficients de synthèse net. On voit que le coefficient varie entre 0,37 g de DBO enlevé/g de SS_v de liqueur mixte à une valeur de 0,72 g de DBO/g de SS_v. La figure 8 montre graphiquement cette observation aussi en fonction du concept de surface de réponse. La figure 8 révèle que le coefficient de synthèse net n'est pas constant et qu'il tend à diminuer sous l'effet combiné d'une augmentation d'âge de boue et de temps de rétention hydraulique. Le coefficient a été calculé sur une base de DBO_{5t} - DBO_{5s}. La différence entre la valeur du coefficient à la condition de 5 heures et 10 jours et celle de 15 heures et 45 jours est de 0,72 à 0,37, soit un facteur de 1,9. Cette variation influence inévitablement le dimensionnement des équipements lorsque le coefficient est utilisé à cette fin. La conception finale de la station s'est arrêtée sur un temps de rétention hydraulique de 12 heures et un âge de boue de 20 à 30 jours. On observera que ces conditions sont dans le faisceau.

Ce choix permettrait une certaine flexibilité car, à 12 heures de rétention hydraulique, le contrôle de l'âge de boue pourrait se situer entre 15 et 25 jours tout en résultant en un indice de volume de boue (mesure de décantabilité) raisonnable. Cette flexibilité pourrait diminuer

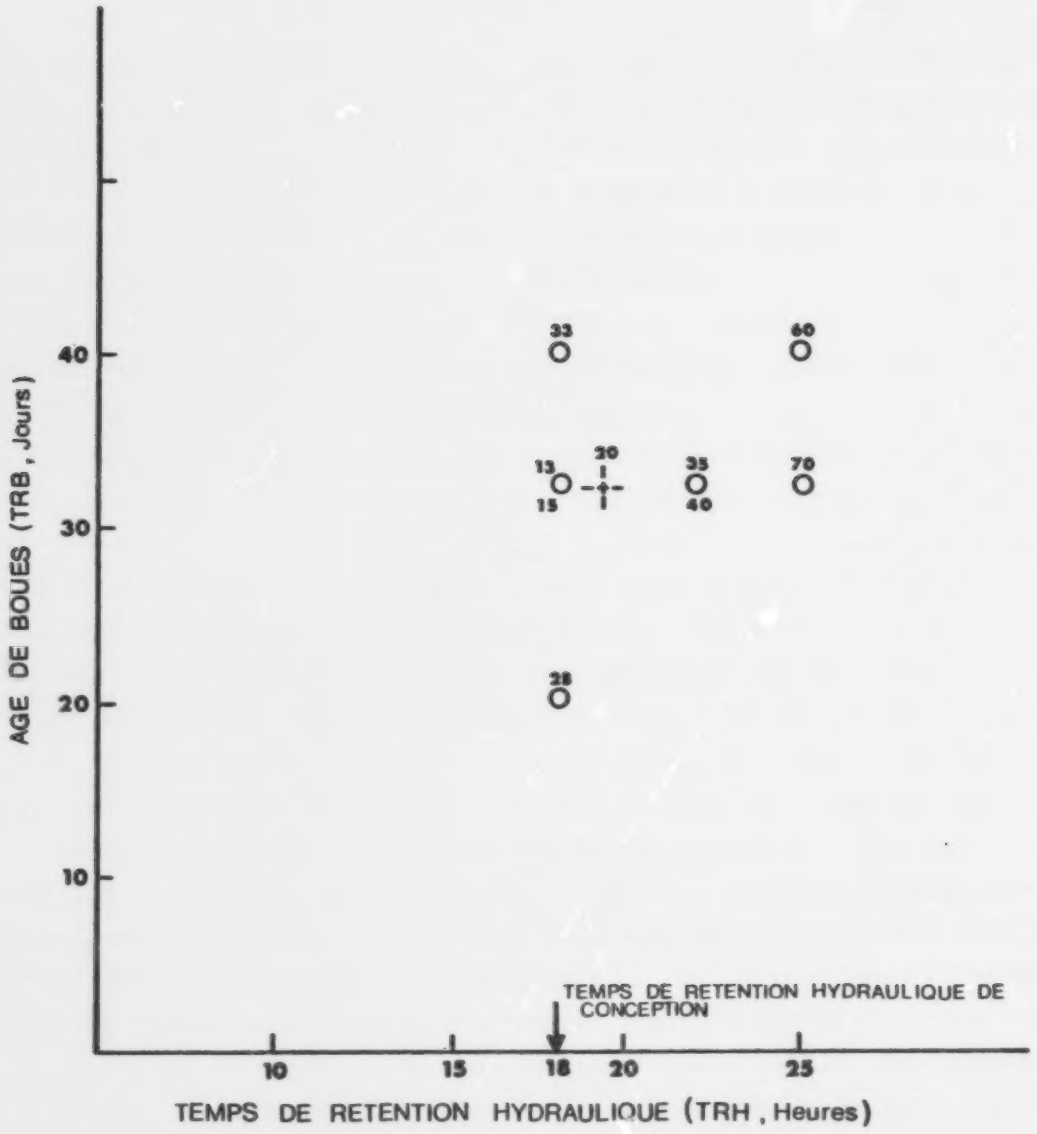
sensiblement si le débit devait augmenter de telle sorte que le temps de rétention soit plus bas que 10 heures. Si ces conditions se produisent, la décantabilité des boues risque fort bien d'être dangereusement compromise.

2. Étude de traitabilité des effluents du procédé de finition de tissus de Dominion Textile.

Cette étude a été exécutée sur les eaux usées d'une usine de finition de tissus. La description du travail a été déjà présentée ailleurs (8). Brièvement, l'objectif de l'étude était de trouver un mode d'opération auquel l'enlèvement de la matière organique et des solides en suspension serait le meilleur, tenant compte du temps de rétention hydraulique auquel la station d'épuration opérait. La figure 9 montre les résultats des solides en suspension dans l'effluent ainsi que les valeurs de réponses sur le plan xy d'une surface. Les reprises ont été exécutées à des périodes différentes. Ces résultats montrent qu'il y a nettement une tendance vers un optimum qui se situerait autour de 18 heures de temps de rétention hydraulique et 32 jours d'âge de boue. On voit qu'en tenant le temps de rétention hydraulique fixe et en augmentant l'âge de boue, on passe par un rendement optimal. Aussi, en maintenant l'âge de boue constant et en diminuant le temps de rétention hydraulique, on se dirige vers cet optimum. Il est probable que le fait de diminuer le temps de rétention hydraulique aurait éventuellement produit un effluent de plus en plus dégradé. Ainsi, ce concept de surface aura permis de localiser rapidement les conditions optimales. L'implantation de cette stratégie à l'échelle a résulté en un effluent ayant 20 mg/l de solides volatiles en suspension (8).

LEGENDE : O Points experimentaux pilotes
-⊕- Point d'opération à l'échelle

Réponse : Solides volatiles en suspension, Mg/L



GRAPHIQUE 9 Représentation des solides volatiles en suspension de l'effluent en fonction d une surface de réponse

Tableau 2

Résultats d'opération de l'étude pilote sur les effluents du procédé de finition des tissus

Âge de boue, d	32	40	32	32	32	21	40	32	30
Temps de rétention hydraulique, h	25	25	22	22	18	18	18	18	18
Solides en suspension dans l'effluent, mg/l	70	60	35	40	13	28	33	15	17
DCO affluent, mg/l	660	759	640	856	845	783	818	680	983
MLVSS, g/l	2,7	3,4	2,7	4,6	4,4	3,3	4,6	3,9	5,8

Comme l'indiquent le tableau 2 et le graphique 9, les reprises effectuées durant cette étude ont donné sensiblement les mêmes résultats malgré des variations importantes de la concentration de l'affluent. Dans les deux cas de reprises, la liqueur mixte s'est ajustée en fonction de la charge organique. Le concept de surface de réponse tient compte de la charge organique quelle que soit sa variation. Dans ce cas, le débit était constant et la concentration a varié. Les coordonnées de la réponse restent les mêmes.

Si la concentration n'avait pas changé et si le débit variait, la charge organique aurait varié proportionnellement et la liqueur mixte se serait établie à un nouvel équilibre. Ainsi, au même âge de boue, les coordonnées de la réponse se déplacent vers une autre valeur de temps de rétention hydraulique.

CONCLUSION

Le concept de réponse présenté dans cette étude offre une façon cohérente d'interpréter les résultats de traitement. Il montre clairement l'effet du temps de rétention hydraulique et de l'âge des boues sur le rendement et les variations du coefficient de synthèse net. Quoiqu'en apparence, le concept de charge organique semble absent, en réalité il est sous-jacent.

Ce concept offre à l'utilisateur un moyen d'incorporer des éléments de flexibilité au niveau de la conception afin de prévenir les effets de changements qualitatifs ou quantitatifs de l'eau usée.

Un tel concept est valable lorsque les conditions d'opération anticipées ou actuelles sont à l'équilibre. Pour permettre une interprétation raisonnable, il est nécessaire d'obtenir un minimum de données expérimentales. La représentation au moyen d'une surface de la performance d'un système biologique met en évidence l'effet de l'âge de boue et du temps de rétention hydraulique. Ainsi, on peut dire qu'à des combinaisons de valeurs faibles ou élevées de ces deux (2) paramètres, la qualité de l'effluent en matières en suspension tendrait à se dégrader. Des combinaisons intermédiaires donneraient des rendements correspondant jusqu'à l'atteinte d'un optimum. La notion de coupe d'iso-réponse constitue l'élément de flexibilité par excellence pour une optimisation économique et technique.

Ce concept n'est pas un modèle cinétique prédictif. Si, par contre, il était possible de développer les équations qui forment la surface pour chaque réponse d'intérêt, tel que la DCO ou la DBO et les solides en suspension, on pourrait obtenir des valeurs réalistes des coefficients nets de synthèse et de la concentration de la liqueur mixte à des conditions différentes de celles

du programme expérimental. Fort de ces données, une conception serait plus fidèlement respectée durant l'opération de la station.

Malgré que seulement deux (2) cas types ont été présentés ici, les notions de ce concept ont été mises à l'épreuve dans une étude sur la traitabilité d'effluents d'une aluminerie et guident présentement l'orientation d'une étude sur le fonctionnement de réacteurs séquentiels dans le secteur agro-alimentaire.

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ACTIVATED SLUDGE FINAL SETTLING TANK ANALYSIS[1]

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INTRODUCTION

Although the biological phase of the activated sludge process is fascinating and has attracted much attention, still effluent quality is established primarily by the solids separation phase. Biological uptake of degradable organic compounds does not ordinarily limit the performance of the activated sludge process; instead, performance is established by the efficacy with which suspended solids synthesized as a consequence of biological removal of degradable organic compounds can be removed. Furthermore, the performance of the solids separation phase of the activated sludge process establishes the concentration of microorganisms that can be maintained in the biological portion of the process (and, hence,its capacity). In addition, the cost of managing waste biological sludge is influenced by the concentration at which excess suspended solids are wasted (and, thus, by the performance of the secondary clarifier).

Activated solids escape separation in secondary settling tanks for two major reasons. The first is the current inability to design settling tanks to achieve the clarification performance they theoretically are capable of (Dick, 1982). The lack of development and implementation of techniques to adequately design inlet, outlet, and flow distribution and control facilities is responsible. The hydraulic limitations account for the common occurrence of moderate concentrations of suspended solids (and, hence, biochemical oxygen demand) in the effluent from activated sludge plants.

The second cause for loss of suspended solids from activated sludge wastewater treatment plants is the application of solids to final settling tanks at rates higher than they can be conveyed to the bottom of the tank. This cause for solids loss can account for massive concentrations of suspended solids in the effluent from activated sludge wastewater treatment plants. Limitations in performance associated with solids overloading also account for inability to maintain adequate concentrations of microorganisms in aeration tanks and for the dilute concentration at which activated sludge solids often are wasted from the process.

This paper concerns the second cause of malperformance of final settling tanks, solids overloading. The solids loading, however, is intimately linked to the biological phase of the activated sludge process. Thus, that linkage and the rational overall design and operation of the biological reactor-solids separator are considered.

The performance of final settling tanks is established by the physical characteristics of the activated sludge solids and by the manner in which the tank is designed and operated. Progress is being made in learning to control the physical characteristics of activated sludge [see, for example, Palm et al., (1980)]; it is to be anticipated that, in time, it will become possible

[1]Presented at Seventh Symposium on Wastewater Treatment, Montreal, Canada, November 21, 1984.

to regulate the settleability of activated sludge. Capability for designing and operating the activated sludge final settling tank so as to obtain best overall performance of the activated sludge process for a sludge with any settling characteristics is available now, however, and review of concepts for optimal final settling tank performance is the focus of this paper.

FACTORS GOVERNING SECONDARY SETTLING TANK PERFORMANCE

The rate at which suspended solids can be transmitted downward in a secondary settling tank by gravity sedimentation and bulk downward movement due to concentrated sludge removal from the bottom of the tank is

$$G = cv + cu \tag{1}$$

where:

G = total suspended solids flux, $M/L^2.T$,
c = suspended solids concentration, M/L^3,
v = sedimentation velocity of sludge solids at concentration c, L/T, and
u = downward velocity caused by sludge removal, L/T.

With uniform withdrawal of thickened sludge over the cross sectional area of the tank, A,

$$u = \frac{R}{A} \tag{2}$$

where R is the rate of recycle of thickened activated sludge to the aeration tank. If, as is common, excess activated sludge is wasted from the secondary clarifier underflow, then R in Equation 2 is replaced by the total thickened sludge underflow rate, Q_u, where

$$Q_u = R + Q_w \tag{3}$$

and Q_w is the rate of sludge wastage. As a practical matter, Q_w is negligible. Consider, for example, an activated sludge process with an 8d mean cell residence time and a 4h actual tank hydraulic residence time (based on the sum of wastewater and recycle flow rates). The total daily mass of solids entering the secondary settling tank then is 6 times the mass of solids in the aeration tank, and, if suspended solids in the secondary tank are excluded from the mean cell residence time calculation, 1/8 of the mass of suspended solids in the aeration tank are wasted daily. The error in ignoring the additional downward velocity caused by Q_w in this case, thus, is about one in 48 or roughly two percent. The error in ignoring waste sludge flow is even less if suspended solids in the secondary tank are included in calculating the mean cell residence time. While the analysis presented here could be modified to include consideration of the waste sludge flow rate, the refinement usually is not justified, and Q_w is ignored throughout this paper.

Experimental evaluation of the relationship between the settling velocity (v) and suspended solids concentration (c) for a particular sludge

reveals that, for a given operating condition (that is, for a given value of the underflow velocity, u), the maximum rate at which solids are able to pass to the bottom of a secondary settling tank is restricted by some particular concentration (the "limiting" concentration) which has a lower value of G in Equation 1 than all other suspended solid concentrations that could exist in the thickening portion of the tank. [See Dick (1972) and similar sources for discussion of solids flux restrictions in continuous settling basins.]

The concentration of suspended solids that can be maintained in the underflow from a settling basin is established by the rate at which suspended solids can be transported downward and by the rate at which sludge is removed from the bottom of the tank. When a secondary clarifier is operated under fully loaded conditions, the underflow suspended solids concentration is

$$c_u = \frac{G_L A}{R} = \frac{c_L v_L + c_L u}{u} = c_L \frac{v_L}{u} + c_L \tag{4}$$

where G_L is the limiting flux (the minimum value of Equation 1 for all suspended solids concentrations that can exist in the thickening portion of the secondary clarifier), c_L is the suspended solids concentration of the rate limiting layer, and v_L is the settling velocity of sludge solids at that concentration. From a mass balance, the concentration of recycled sludge also can be written as

$$c_u = \frac{(Q_o + R)\, X - Q_o c_e}{R} \tag{5}$$

where Q_o is the wastewater flow rate, X is the mixed liquor suspended solids concentration, and c_e is the effluent suspended solids concentration. Ordinarily, the mass of suspended solids in the effluent ($Q_o c_e$) is a negligible part of the suspended solids loading on the tank [$(Q_o + R)\, X$], and

$$c_u = \frac{(Q_o + R)\, X}{R} \tag{6}$$

[Ordinarily it is quite tolerable to ignore suspended solids lost in the effluent. If, for example, c_e is 20 mg/L, X is 3000 mg/L, and R is 50 percent of Q_o, then $Q_e c_e$ represents only about 0.4 percent of the total mass of suspended solids applied to the secondary settling tank. The technique for secondary settling tank analysis presented in this paper could be modified to avoid the assumption of negligible suspended solids loss in the effluent, but such refinement usually is not warranted.]

If the applied areal solids loading [$(Q_o + R)\, X/A$] exceeds the

limiting solids flux ($c_L v_L + c_L u$), solids overloading occurs. The excess of the applied loading over the limiting flux will accumulate at the concentration of the rate limiting layer and await opportunity to be transmitted downward through the bottleneck presented by the rate limiting layer. If the solids overloading is relieved before the rate limiting layer propogates upward to the elevation of the inlet, then clarification failure does not occur; the inventory of solids in the settling tank merely increases (to be withdrawn later during a period of solids underloading). If, however, solids overloading continues beyond the time that the rate limiting layer reaches the inlet elevation, then a mass balance dictates that solids move upward into the clarification zone [at a rate equal to $(Q_o + R)\ X/A - (c_L v_L + c_L u)$]. Under these conditions, the suspended solids concentration in the clarification zone, c_c, can be found by equating the solids overload to the total upward flux caused by upward fluid flow less the flux caused by particle sedimentation to give

$$c_c = \frac{(Q_o + R)\ X/A - (c_L v_L + c_L u)}{Q_o/A - v_c} \tag{7}$$

where v_c in the gravity sedimentation velocity of activated sludge solids at concentration c_c. As solids overloading continues a blanket at concentration c_c will steadily move upward (at velocity $Q_o/A - v_c$) until it reaches the effluent weir. From a mass balance, the effluent suspended solids concentration, c_e will be

$$c_e = \frac{(Q_o + R)\ X - (c_L v_L - c_L u)A}{Q_o} \tag{8}$$

[Note that, because sedimentation occurs in the clarification portion of a secondary tank (at rate v_c), $c_e < c_c$. Under sustained solids overloading, however, sedimentation in the clarification portion of the secondary tank is of no consequence, and the effluent suspended solids concentration is established solely by the mass balance shown in Equation 8.]

The relentless upward propagation of activated sludge solids past the inlet of overloaded secondary tanks and on to the effluent weir is a regretfully frequent occurrence. It commonly is referred to as "bulking sludge," and blamed on poor settleability. While adverse settleability contributes to effluent quality deterioration attributable to solids overloading, it is but one variable (v_L) amongst the many in Equation 8.

INFORMATION REQUIRED TO ANALYZE PROCESS PERFORMANCE

For a given wastewater, two principal factors control performance of activated sludge facilities. The first is the concentration of microrganisms that can be maintained in the aeration tank. This parameter establishes the maximum capacity of the biological reactor for removing biochemical oxygen demand. [see Lawrence and McCarty (1970) for consideration

of the relationship between the population of microorganisms maintained in a biological treatment process and process performance.] The second factor that establishes activated sludge process performance is the efficacy with which suspended solids are separated by the final settling tank. [Both the biochemical oxygen demand and the suspended solids content of activated sludge process effluent are established by the effectiveness of solids separation - see Dick (1970) for discussion of the significance of solids separation on process performance.]

Operational control of the two major factors affecting activated sludge process performance (mixed liquor suspended solids concentration and effluent suspended solids concentration) can be maintained by simultaneously assessing three variables that establish process performance. These variables are: (1) the solids loading on the secondary settling tank, (2) the manner in which the secondary settling tank is operated, and (3) the sedimentation characteristics of the activated sludge solids. The ways in which these three factors influence activated sludge process performance are considered in the sections that follow.

Solids Loading on Secondary Settling Tank

The applied solids loading, G_a, on a secondary settling tank with horizontal cross sectional area, A, is

$$G_a = \frac{(Q_o + R)X}{A} \tag{9}$$

where Q_o is the wastewater flow rate, R is the recycle flow rate, and X is the mixed liquor suspended solids concentration at the point of discharge from the aeration tank. The applied solids loading or applied flux is, thus, the areal rate of solids application with units $M/L^2{}_vT$ (for example, $kg/m^2 \cdot d$).

For purposes of analysis it is expedient to consider separately the suspended solids conveyed to the final settling tank by wastewater flow and those conveyed by recycled flow. That is, it is convenient to write Equation 9 as

$$G_a = \frac{Q_o X}{A} + \frac{RX}{A} \tag{10}$$

and to consider independently the two terms on the right side of the equation. The first term on the right side of Equation 10 can be depicted graphically as shown in Figure 1. The slope of the line is the traditional secondary clarifier overflow rate, Q_o /A. The line shows applied solids flux corresponding to any mixed liquor suspended solids concentration, and is called the loading line.

Figure 2 illustrates the addition of the flux of solids conveyed by the recycled flow to that conveyed by wastewater flow to obtain the total applied solids flux. For convenience of analysis, however, it is useful to depict

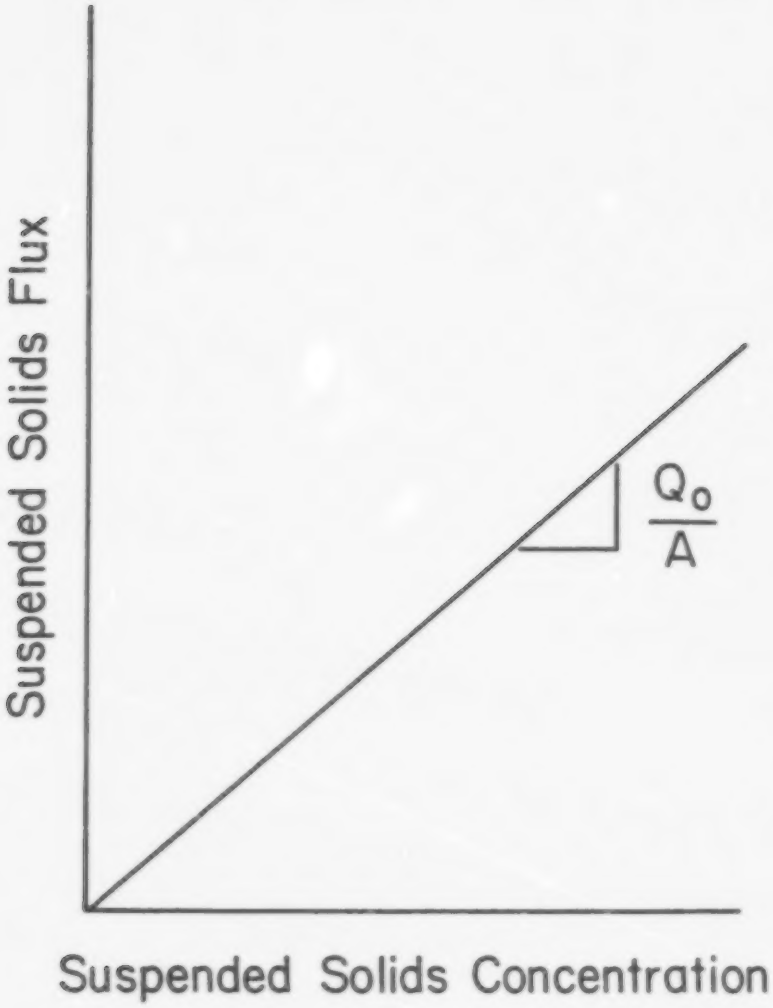

FIGURE 1. THE APPLIED SOLIDS FLUX CONVEYED TO THE FINAL SETTLING TANK BY WASTEWATER FLOW CAN BE GRAPHICALLY PORTRAYED AS ILLUSTRATED HERE. THIS LINE, THE LOADING LINE, IS ONE OF THREE REQUIRED FOR ANALYSIS OF SEDIMENTATION TANK PERFORMANCE.

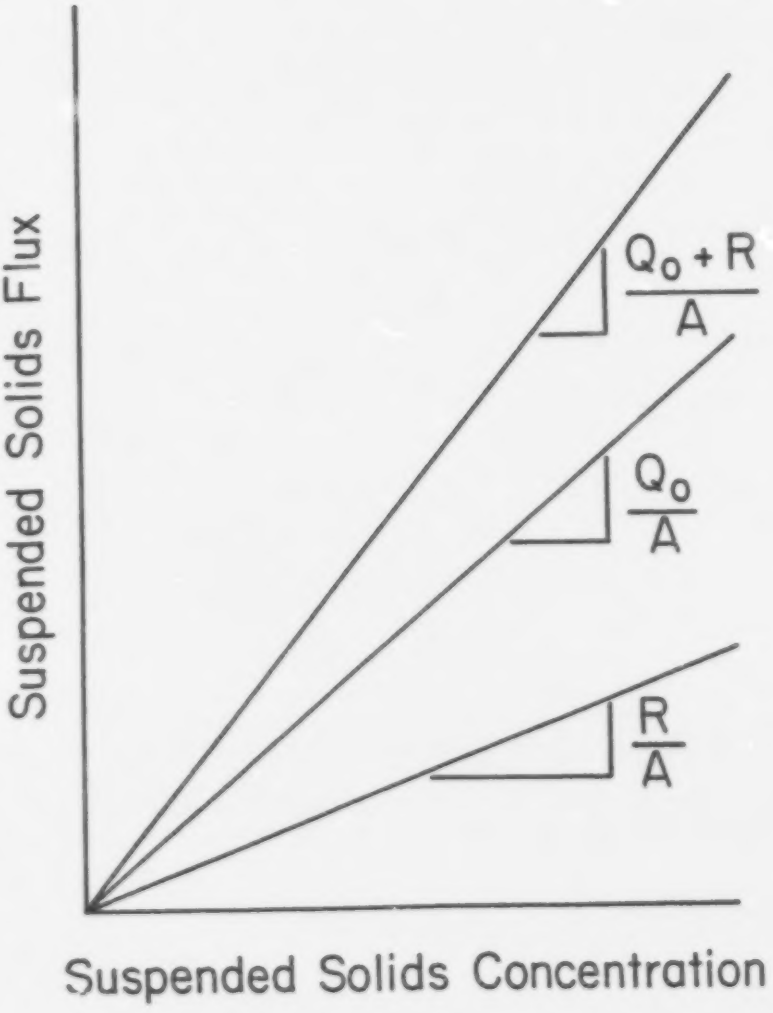

FIGURE 2. THE TOTAL APPLIED SOLIDS FLUX TO THE SECONDARY SETTLING TANK IS THAT CONVEYED BY WASTEWATER FLOW PLUS THAT CONVEYED BY RECYCLE AS INDICATED IN EQUATION 10 AND ILLUSTRATED GRAPHICALLY HERE.

the solids loading conveyed to the secondary settling tank by recycle in a different way as illustrated in the following discussion of the second major factor influencing activated sludge process performance.

Secondary Clarifier Operating Conditions

To analyze performance of the activated sludge process, it is necessary to characterize the manner in which the secondary settling tank is operated. The principal secondary settling tank operational variable is the rate of removal of sludge from the bottom of the tank. The rate of sludge recycle influences the secondary tank loading (Equation 9), solids handling capacity (Equations 1 and 2), and underflow suspended solids concentration (Equation 6). When a secondary clarifier is operated at or below the limiting solids handling capacity,

$$Rc_u = GA \tag{11}$$

Combining Equations 2 and 11 yields

$$u = \frac{G}{c_u} \tag{12}$$

Thus, as illustrated in Figure 3, the slope of a line connecting the solids flux, G, through a secondary settling tank and the suspended solids concentration of the sludge withdrawn from the bottom of tank is the downward velocity, u, caused by removal of sludge from the bottom of the tank. The line is called the "operating line" because it defines the way the settling tank is operated.

The operating line also can be used to graphically demonstrate the solids loading conveyed to the secondary settling tank by recycled flow. This is because the slope of the operating line is u, and

$$u = \frac{R}{A} \tag{2}$$

Thus the operating line conveys the same information as the line in Figure 2 showing the suspended solids loading conveyed to the secondary settling tank by the recycled flow, but it has opposite slope.

Activated Sludge Settling Characteristics

The third item of information required to assess secondary settling tank operation is a characterization of the settleability of suspended solids in the sludge. This can be done by measuring the liquid-solid interface settling velocity at several different suspended solids concentrations so as to develop the concentration-velocity relationship and then to plot a curve of solid flux due to sedimentation as a function of suspended solids concentra-

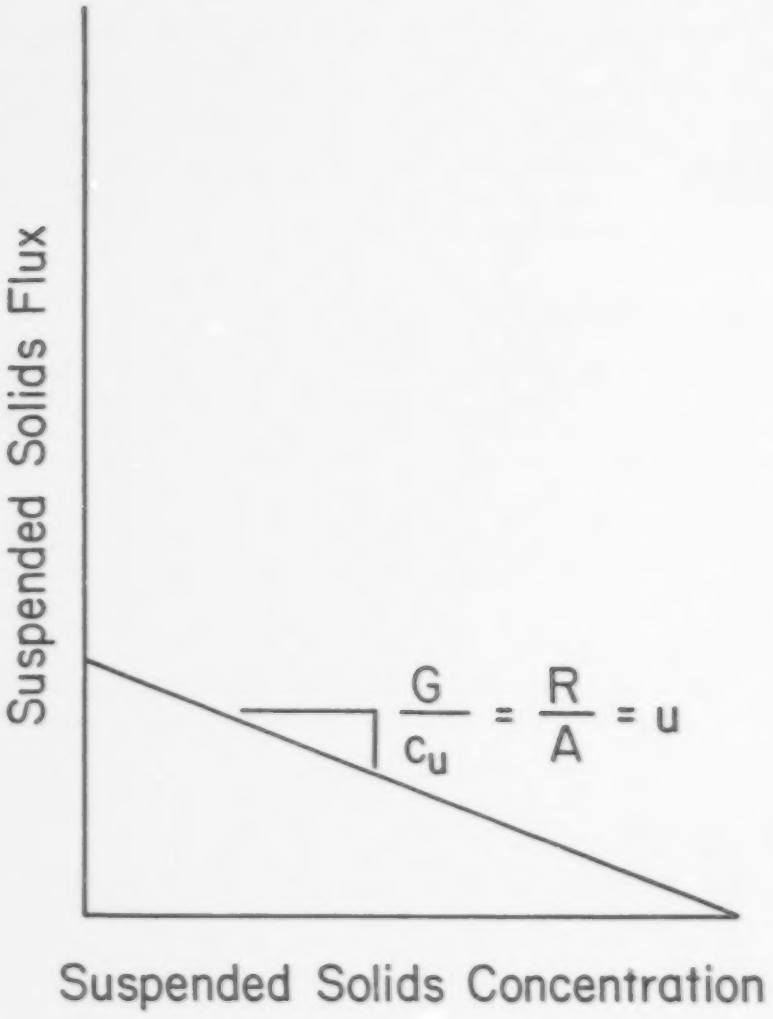

FIGURE 3. THE SECOND LINE REQUIRED FOR ANALYSIS OF THE ACTIVATED SLUDGE PROCESS IS THE OPERATING LINE. IT IS FORMED BY CONNECTING THE TOTAL SOLIDS FLUX THROUGH A SETTLING TANK WITH THE UNDERFLOW SUSPENDED SOLIDS CONCENTRATION. IT HAS A SLOPE EQUAL TO THE DOWNWARD VELOCITY CAUSED BY REMOVING A SLUDGE FROM THE TANK BOTTOM, AND IT DEPICTS THE WAY IN WHICH A SETTLING TANK IS OPERATED.

tion. This "batch flux curve" can be used to establish the limits in settling tank performance imposed by the settleability of the specific sludge being settled.

A typical batch flux curve for activated sludge is shown in Figure 4. It passes through the origin because c is zero there, approaches the abscissa at high suspended solids concentrations because the settling velocity of compressible suspensions approaches zero at high concentrations, and shows that the maximum capacity for transporting solids by gravity sedimentation occurs at intermediate suspended solids concentrations.

REVELATIONS FROM COMBINING CURVES

Superimposition of the operating line (Figure 3) with the line showing the solids loading conveyed to the settling tank by wastewater flow (Figure 1) provides insight on the interaction between aeration tanks and settling tanks. This is illustrated in Figure 5. The intersection of the two lines is called the statepoint (McHarg, 1973). As seen from Equation 10, the only suspended solids concentration that fulfills Equation 1 is X, the mixed liquor suspended solids concentration. Thus, construction of the lines shown in Figure 5 provides a convenient assessment of how operation of the secondary clarifier affects the biological phase of the activated sludge process, for it identifies the mixed liquor suspended solids concentration corresponding to any recycle rate.

Furthermore, as shown in Figure 5, the intersection of the loading line with the operating line provides graphical representation of the secondary settling tank loading conveyed by the wastewater $(\frac{Q_o}{A} X)$ as opposed to that conveyed by recycle $(\frac{R}{A} X)$.

Combination of the operating line from Figure 3 with the batch flux curve from Figure 4 permits assessment of whether operating conditions are compatible with the sedimentation properties of the activated sludge solids. This is illustrated in Figure 6. When the batch flux curve line is above the operating line, adequate capacity exists to transport suspended solids to the bottom of the secondary clarifier. This is illustrated for concentration c_1 in Figure 6 where it is apparent that

$$G < c_1 v_1 + c_1 u. \tag{13}$$

The dilute suspended solids concentration at which the operating line and the batch flux curve cross in Figure 6 is the only concentration that has solids handling capacity exactly equal to the applied flux (it is the only concentration that satisfies Equation 1), and, hence, it is the only suspended solids concentration that will exist in the portion of the secondary clarifier with full cross-sectional area, A. The operating line is above the batch flux curve to the left of the intersection, but that is of no concern if, as is usual, the mixed liquor suspended solids concentration is to the right of the intersection. Thus, the operating line shown in Figure 6 is successful because final sedimentation tank performance would not be restricted by sludge settleability.

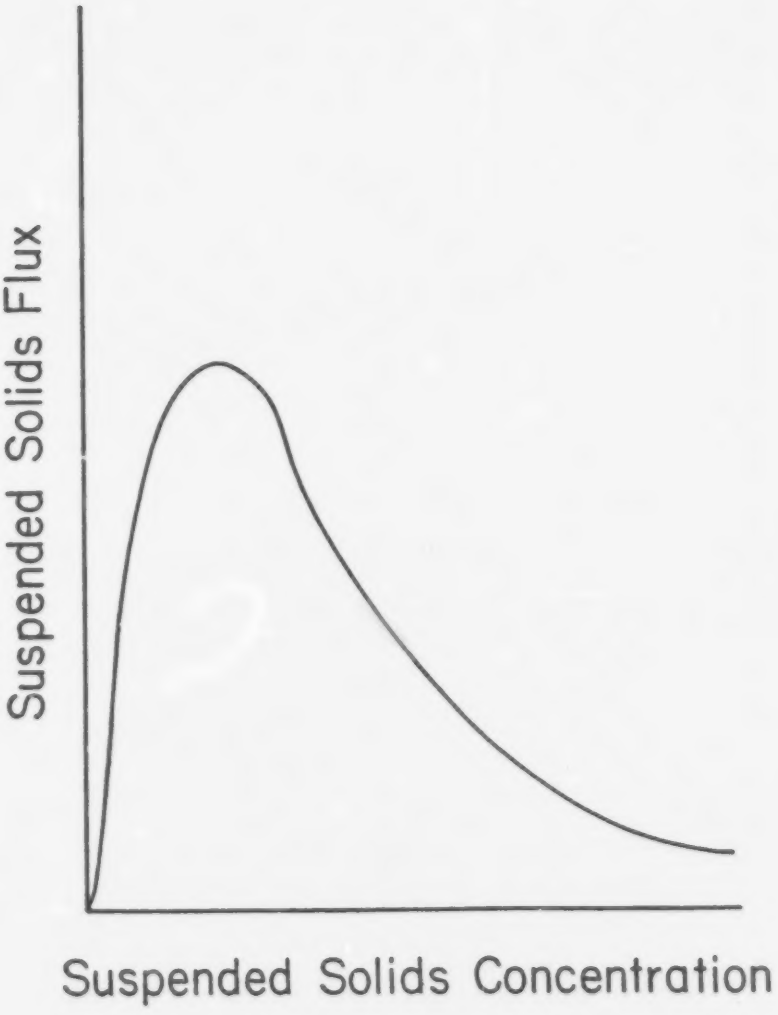

FIGURE 4. THE BATCH FLUX CURVE CHARACTERIZES THE CAPACITY OF A GIVEN ACTIVATED SLUDGE TO CONVEY SUSPENDED SOLIDS TO THE BOTTOM OF A SETTLING TANK BY GRAVITY SEDIMENTATION. IT IS THE LAST OF THE THREE CURVES NEEDED TO ANALYZE ACTIVATED SLUDGE PROCESS PERFORMANCE.

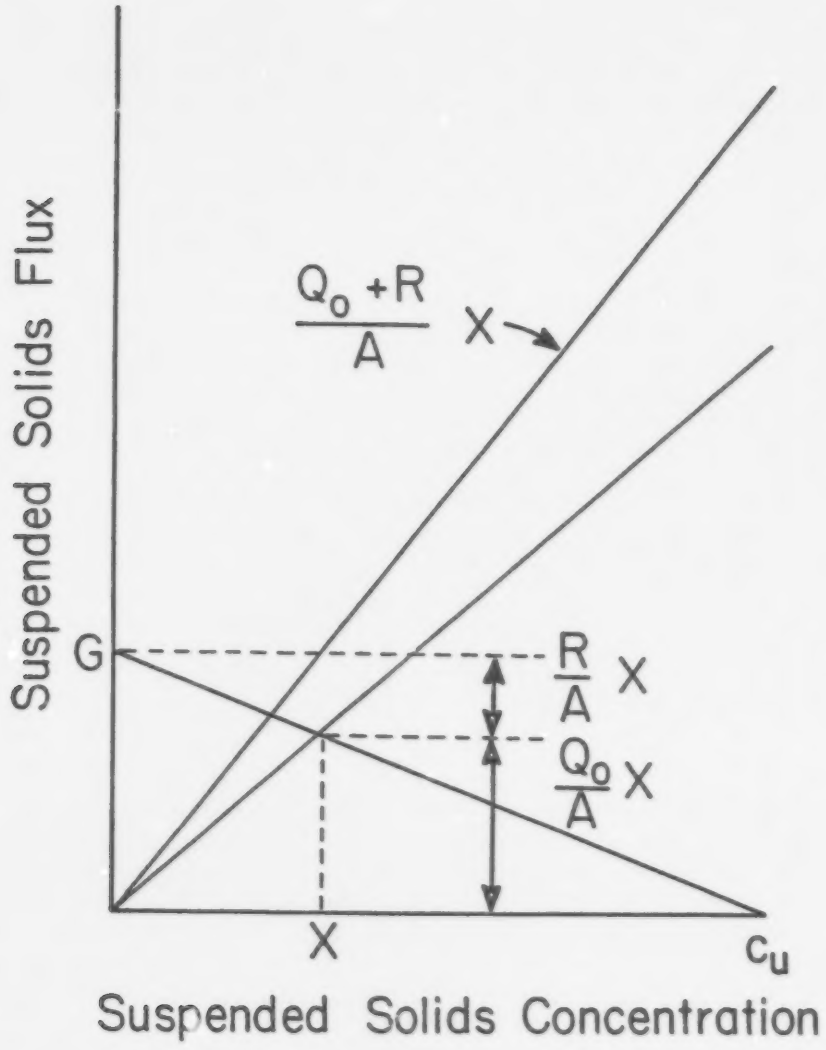

FIGURE 5. THE INTERSECTION OF THE OPERATING LINE WITH THE LINE SHOWING SOLIDS FLUX CONVEYED BY WASTEWATER FLOW, THE STATE POINT, IDENTIFIES THE MIXED LIQUOR SUSPENDED SOLIDS CONCENTRATION.

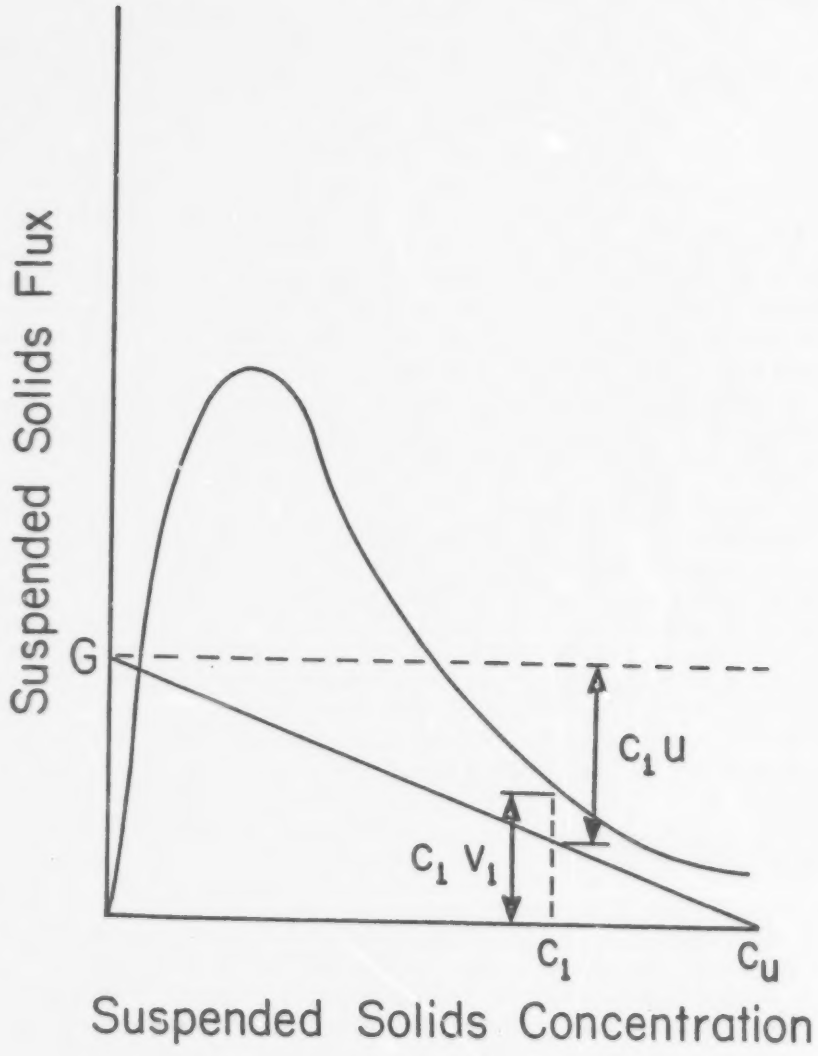

FIGURE 6. WHEN THE BATCH FLUX CURVE LIES ABOVE THE OPERATING LINE, CAPACITY FOR TRANSMITTING SOLIDS TO THE BOTTOM OF A SECONDARY CLARIFIER EXCEEDS THE APPLIED LOADING, AND OPERATION IS SUCCESSFUL.

The consequence of allowing the operating line to be above the batch flux curve is illustrated in Figure 7. It is seen that the secondary clarifier will fail because

$$G' > c_1 v_1 + c_1 u. \qquad (14)$$

Solids applied in excess of the limiting solids handling capacity will be forced to propogate upward and, ultimately, will be contained in the effluent at the concentration given by Equation 8.

It is apparent from Figures 6 and 7 than an operating line that makes full use of the solids handling capacity of secondary clarifiers is one that is tangent to the batch flux curve (as illustrated in Figure 8). The point of tangency identifies the rate limiting concentration, c_L, which, because it fulfills Equation 1, will exist in the full cross-sectional area of the secondary clarifier. If, as in Figure 7, the applied solids flux exceeds the limiting flux, then the secondary clarifier will perform to the extent of its capability (as defined in Figure 8), and solids that constitute "overloading" will be lost in the "clarified" effluent.

Simultaneous consideration of all three of the curves that establish performance of the activated sludge process (Figures 1, 3, and 4) gives a convenient and concise picture of steady state performance of the activated sludge process. This is illustrated in Figure 9. It is seen that, with a wastewater flow rate, Q_o, and sludge settleability as defined by the batch flux curve, selection of recycle rate R and underflow concentration c_u results in successful clarifier performance and gives mixed liquor suspended solids concentration, X. As suggested in the following section, however, the operating line shown may not be the best one.

STRATEGIES FOR DESIGN AND OPERATION

Within limits imposed by sludge settleability, wastewater loading, and facilities provided by the treatment plant designer, the operator of an activated sludge wastewater treatment plant has great flexibility in selecting operating conditions. This is illustrated in Figure 10 where two extreme operating lines compatible with the wastewater flow rate shown in Figure 1 and the activated sludge settleability defined by Figure 4 are illustrated. Operating line 1 makes full use of the installed aeration and clarification facilities; the secondary clarifier is operated at limiting conditions and the mixed liquor suspended solids concentration could not be higher. The operating condition also represents the point of incipient failure, and, hopefully, it is one to which the operator would be forced only by extreme loading conditions. The operating line is probably undesirable for routine operation because it leaves no reserve capacity to accommodate changes, requires a high recycle rate (nearly 160 percent of Q_o), maximizes air consumption (because the mixed liquor suspended solids concentration is high), and might stimulate undesirable organisms associated with low organic loading intensities (Sezgin, et al., 1978).

Operating line 2 makes very poor use of installed capital investment. It would not represent a prudent design condition, but it would be a very comfortable operating line. Much reserve capacity exists in both the

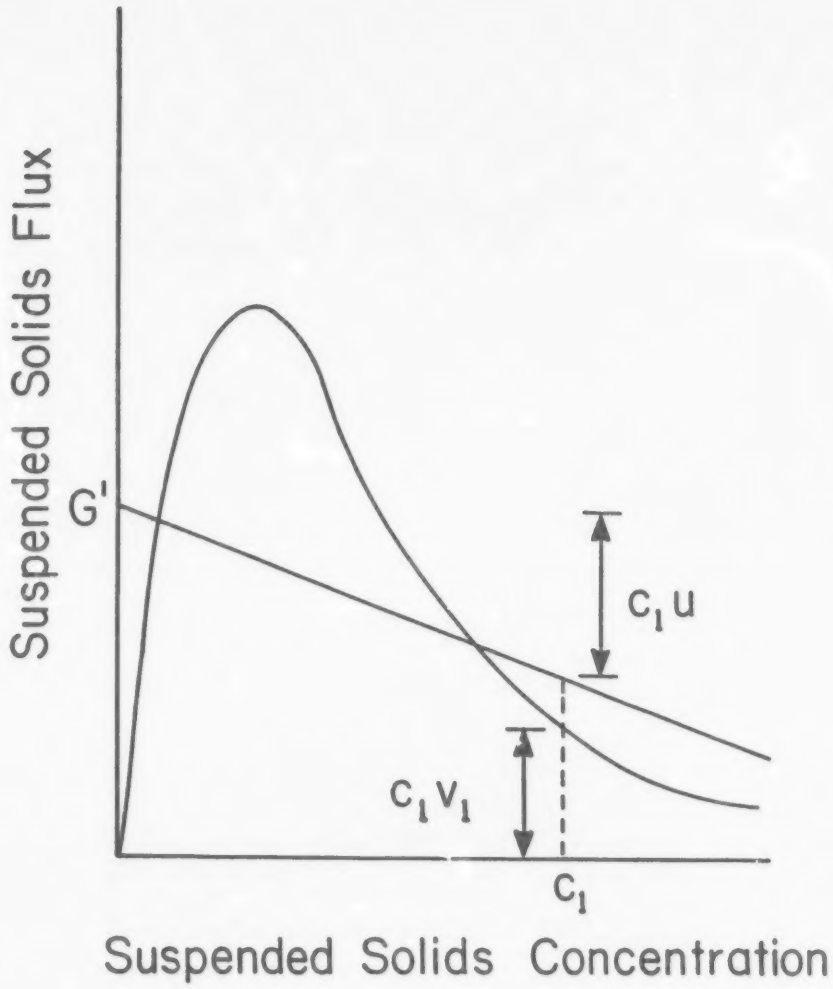

FIGURE 7. WHEN THE OPERATING LINE LIES ABOVE THE BATCH FLUX CURVE FOR SUSTAINED PERIODS OF TIME, FAILURE OCCURS BECAUSE THE CAPACITY FOR TRANSMITTING SOLIDS TO THE BOTTOM OF THE TANK IS LESS THAN THE APPLIED SOLIDS LOADING.

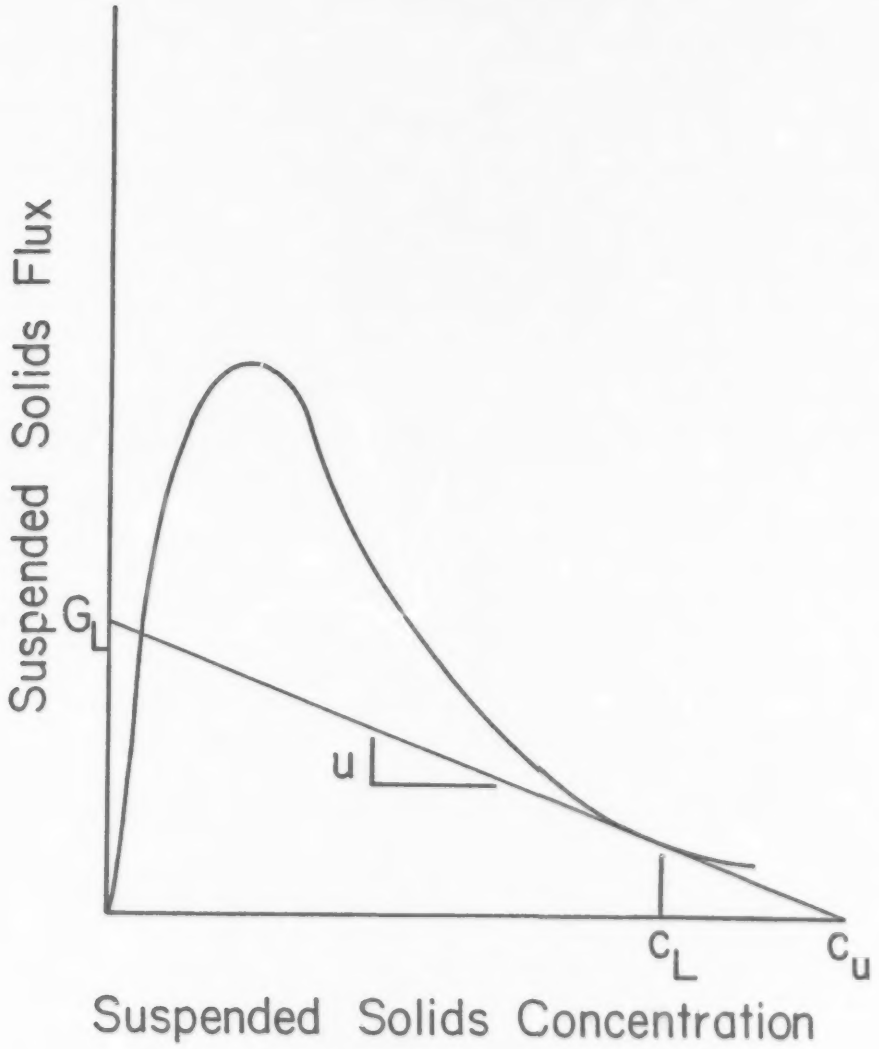

FIGURE 8. SECONDARY CLARIFIERS ARE FULLY LOADED WHEN THE OPERATING LINE IS TANGENT TO THE BATCH FLUX CURVE. FOR THE RECYCLE RATE THAT GIVES THE VALUE OF u SHOWN, SLUDGE AT A CONCENTRATION c_L LIMITS CAPACITY TO G_L. ANY SOLIDS APPLIED IN EXCESS OF G_L WILL BE LOST IN THE EFFLUENT.

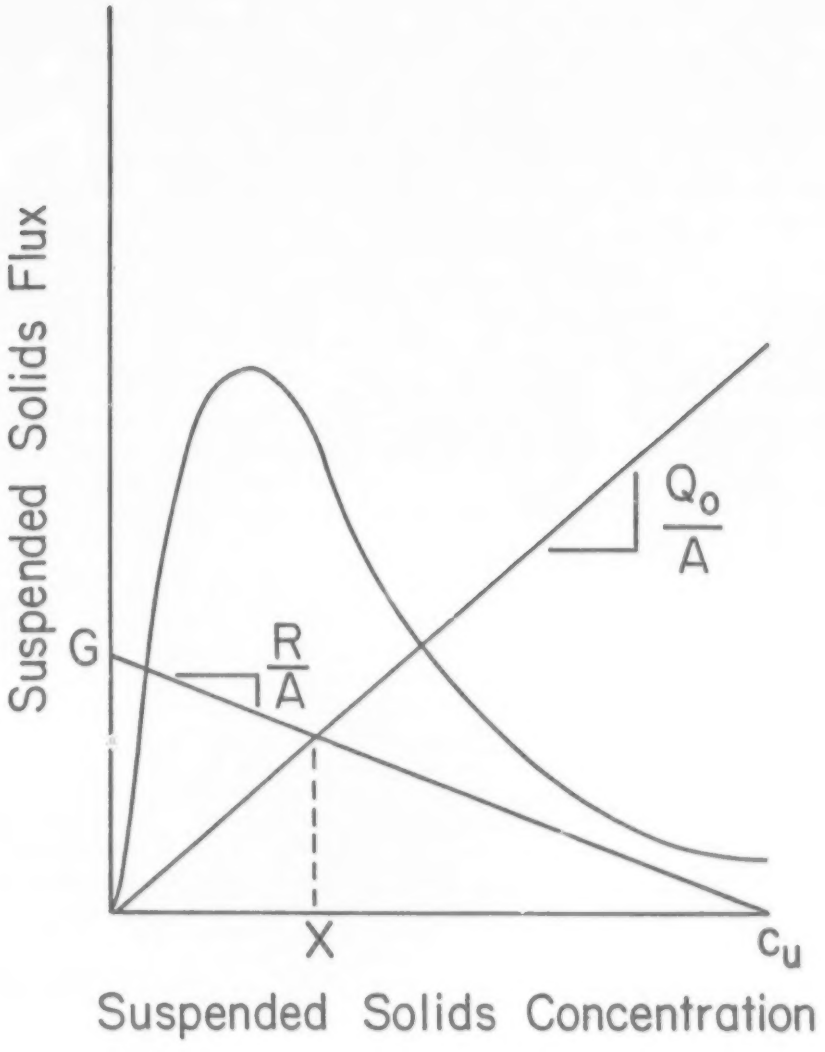

FIGURE 9. FROM OBSERVATION OF THE LOADING LINE, BATCH FLUX CURVE, AND OPERATING LINE, IT CAN BE SEEN THAT THE SECONDARY SETTLING TANK IS NOT OVERLOADED AND THE MIXED LIQUOR SUSPENDED SOLIDS CONCENTRATION IS X.

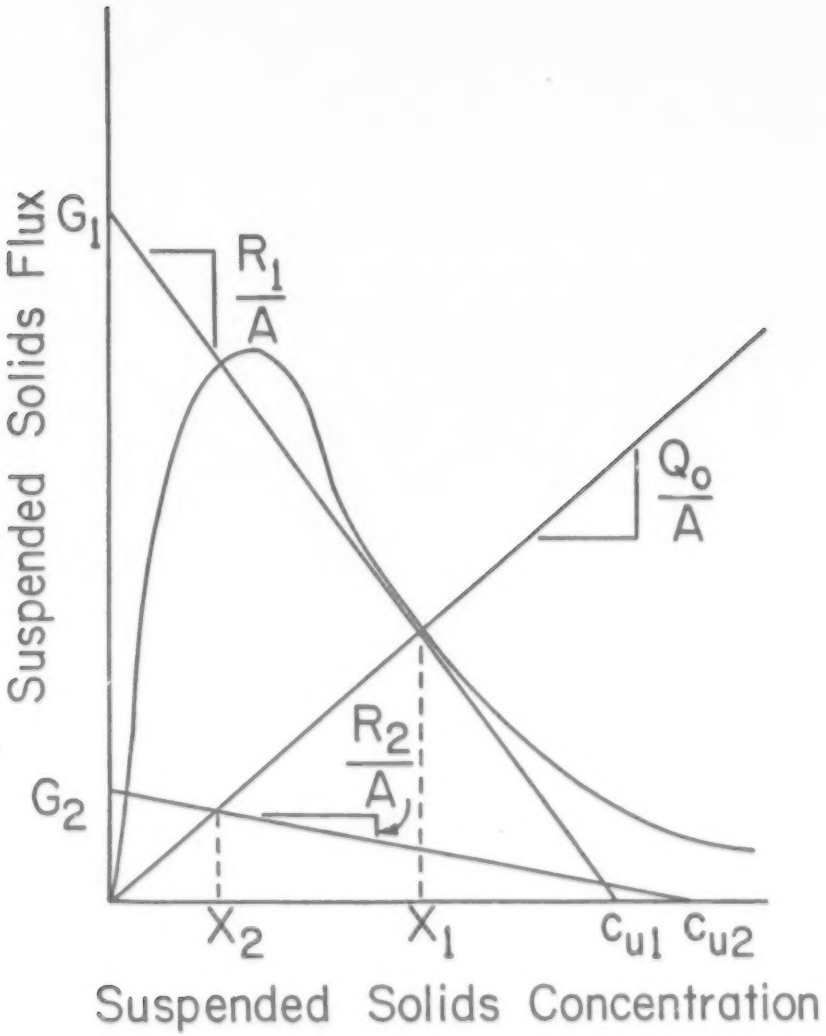

FIGURE 10. OPERATING LINE 1 BOTH MAXIMIZES THE MIXED LIQUOR SUSPENDED SOLIDS CONCENTRATION AND USES THE FULL CAPACITY OF THE SECONDARY CLARIFIER. INSTALLED FACILITIES ARE DRASTICALLY UNDERLOADED WHEN OPERATING LINE 2 IS USED.

aeration tank and clarifier. The recycle rate is only about 20 percent of Q_o, and the low mixed suspended solids concentration minimizes aeration costs. A disadvantage of the second operating line is that the high biological process loading would maximize excess sludge production, and, in extreme cases, ability to remove soluble biochemical oxygen demand could be impaired. Also, undesirable organisms stimulated by the high organic loading intensities could become a problem (Sezgin, et al., 1978).

As a practical matter, the ideal operating line is probably one between the extremes created by conditions 1 and 2. The identity of the best operating line is unique to circumstances at individual plants.

Strategy for dealing with changes in treatment plant loading or activated sludge settleability can be developed from plots showing loading lines, batch flux curves, and operating lines. Use of the three curves to develop strategy for dealing with changes in wastewater flow rate is illustrated in Figure 11. The batch flux curve is the same as in Figure 4, and the loading line, Q_o/A, is from Figure 1. Loading lines representing conditions when the wastewater flow rate is half normal and twice normal are also included. The strategy illustrated in Figure 11 for dealing with different wastewater flow rates is to select conservative operating conditions such that wide variations in wastewater flow can be accommodated without change in the recycle rate. (The slopes of all three operating lines shown in the figure are equal, indicating no change in rate of recycle.) Two different operational strategies are illustrated in the figure, and they differ in the amount of sludge that is wasted from the activated sludge process. One strategy, for which state points are shown in circles, is to maintain a constant mixed liquor suspended solids concentration. The other (with state points in squares) is to increase the mixed liquor suspended solids concentration when reduction in wastewater flow rate allows.

An alternative strategy for dealing with changes in wastewater flow rate is shown in Figure 12. Here, a strategy of maintaining the activated sludge recycle rate as a fixed percentage of the wastewater flow rate is evaluated. It is seen that, for the recycle fraction selected, state points comfortably close to full sedimentation tank loading are achieved.

As an alternative to the operational remedies for wastewater flow rate variations illustrated in Figures 11 and 12, each new loading condition could be considered independently taking into account anticipated future changes, the desire to minimize costs for aeration, waste sludge management, and sludge recycle, as well as other factors that influence operational decisions. This approach is preferable, but it is less routine than the solutions illustrated in Figures 11 and 12 and requires the attention of a skilled operator.

A systematic plan for dealing with changes in wastewater flow rate and changes in activated sludge settleability is suggested in Figure 13. As indicated by the various loading lines, it is considered that wastewater flow varies over a wide range. Also, as indicated by the multiple batch flux curves, it is considered that settleability changes. A possible rational approach to operation of such a plant would be to conduct spot checks of settleability so as to locate an operational space in Figure 13. If, for example, activated sludge settleability was poor and moderate wastewater loading was being experienced, then an operating line appropriate for the area designated as "1" should be selected as indicated by the line with slope R_1/A. Similarly, recycle rate R_2 would be appropriate when settleability and

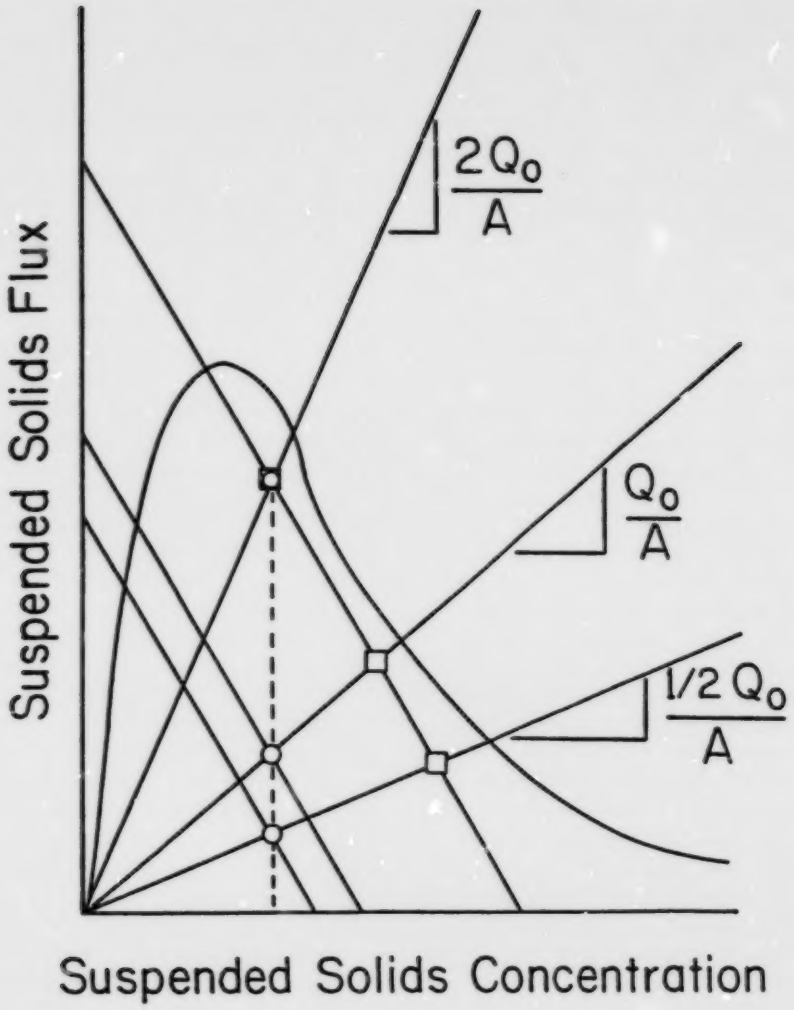

FIGURE 11. STRATEGIES FOR DEALING WITH WASTEWATER FLOW RATE CHANGES CAN BE CONVENIENTLY EVALUATED. THOSE CONSIDERED HERE INVOLVE NO CHANGE IN ACTIVATED SLUDGE RECYCLE RATE.

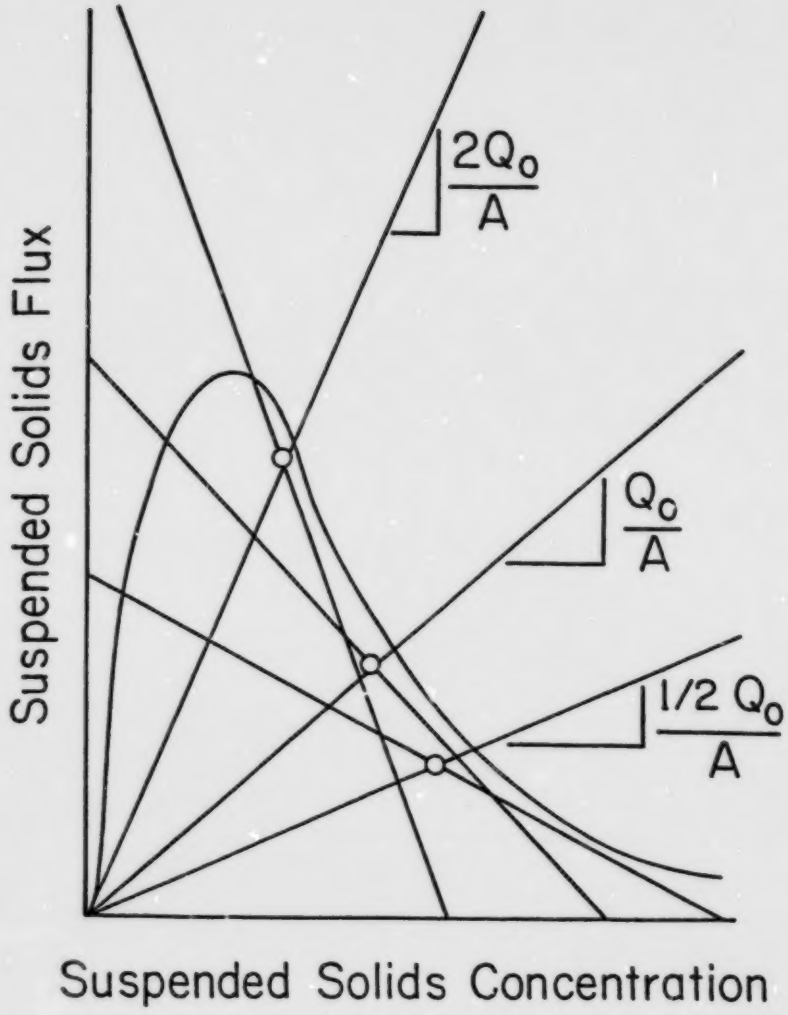

FIGURE 12. USE OF A FIXED RECYCLE RATIO WOULD GIVE THE STATE POINTS INDICATED WHEN WASTEWATER FLOW RATE CHANGED.

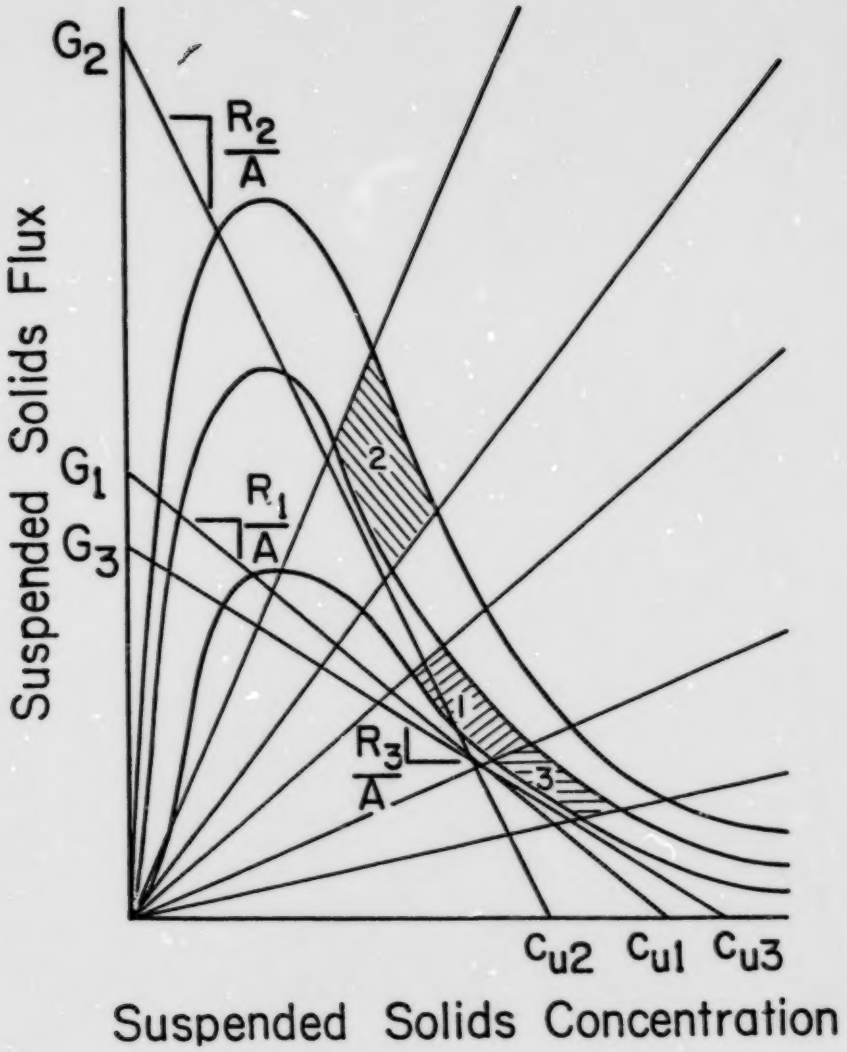

FIGURE 13. IT MAY BE EXPEDIENT TO DIVIDE THE OPERATIONAL DOMAIN INTO SECTORS DEPENDING ON WASTEWATER FLOW RATE AND ACTIVATED SLUDGE SETTLEABILITY AND TO PRESCRIBE APPROPRIATE OPERATING CONDITIONS FOR EACH SECTOR.

wastewater flow rate were in Region 2, and recycle rate R_3 would be called for when conditions fell in Region 3. A prescribed technique such as suggested in Figure 13 would allow unskilled operating personnel to respond to changes in loading and settleability in a conservative fashion that would maximize use of facilities while avoiding the chance of solids loss due to secondary clarifier overloading. More finely tuned operational strategy could be obtained by avoiding the "zones" indicated in the figure and considering each operating condition in the light of all other factors influencing the choice of operating conditions.

NONSTEADY STATE CONDITIONS

All operating conditions considered in this paper are based on steady-state conditions. Transient conditions imposed, for example, by sudden changes in wastewater flow rate could result in loss of solids in the effluent. These nonsteady state perturbations are not considered here (but see, for example, Bisogni and Dick, 1978). The practice advocated in this paper of avoiding steady-state operation at fully loaded conditions minimizes the possibility for solids loss during nonsteady state conditions because avoidance of operation with a rate limiting layer in the secondary sedimentation tank maximizes the storage volume for solids that reach secondary sedimentation tanks during temporary periods of overloading.

SUMMARY

The performance of most activated sludge wastewater treatment plants is established by the success with which activated sludge solids are removed from the process effluent. Hydraulic imperfections in sedimentation tanks account for small concentrations of activated sludge solids in effluents, but large losses of suspended solids ordinarily are attributable to secondary clarifier suspended solids loadings in excess of the rate at which solids can be transported downward in the sedimentation tank.

Rational assessment of conditions influencing secondary clarifier performance allows identification of operating conditions that make full use of installed aeration and clarification capacity while avoiding loss of solids in the effluent due to solids overloading. A convenient graphical procedure makes use of three curves; (1) a "loading line" that indicates the amount of activated sludge solids carried to the secondary sedimentation tank by wastewater flow, (2) an "operating line" that indicates the underflow suspended solids concentration, recycle rate, and solids loading and (3) a "batch flux curve" that demonstrates the capacity of the activated sludge for transmitting solids through the final sedimentation tank by gravity settling.

Simple operational rules for controlling activated sludge processes based on rational interpretation of secondary clarifier performance are considered. More precise control could be obtained by considering the three factors influencing performance outlined in the previous paragraph in the context of other variables influencing optimal activated sludge process performance.

NOMENCLATURE

Symbols

A = Horizontal cross sectional area of sedimentation tank, L^2
c = Suspended solids concentration, M/L^3
G = Suspended solids flux, $M/L^2 \cdot T$
Q = Flow rate, L^3/T
R = Rate of recycle of settling tank underflow to the aeration tank, L^3/T
u = Downward velocity of suspension in settling tank caused by thickened sludge withdrawal, L/T
v = Sedimentation velocity, L/T
X = Mixed liquor suspended solids concentration, M/L^3

Subscripts

a = Applied
c = Clarification zone
e = Effluent
L = Layer that limits the suspended solids flux, G
o = Wastewater
u = Sedimentation tank underflow
w = Waste activated sludge

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A NEW DESIGN PROCEDURE FOR ACTIVATED SLUDGE BASED ON ACTIVE MASS

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INTRODUCTION

This paper will present a concept of biological waste treatment design based on degradable fraction of the mixed liquor. Early design equations equated the active biomass to the mixed liquor volatile suspended solids. (1) These equations have been modified by Eckenfelder (2), to acknowledge the role of degradable biomass as an indicator of the viable fraction of a systems' biomass. Eckenfelder next developed an empherical relationship between organic loading and degradable fraction (2) and incorporated a correction factor into its basic design equations. In this paper material balances are employed to relate degradable fraction to sludge age. The resulting relationships are then employed to define sludge yield and a modified kinetic relationship for process design.

Determination and Definition of Degradable Fraction

The degradable fraction of an activated sludge mixed liquor is determined by a batch endogenous respiration study. The endogenation of mixed liquor volatile suspended solids follows first order kinetics and is illustrated in Figure 1.

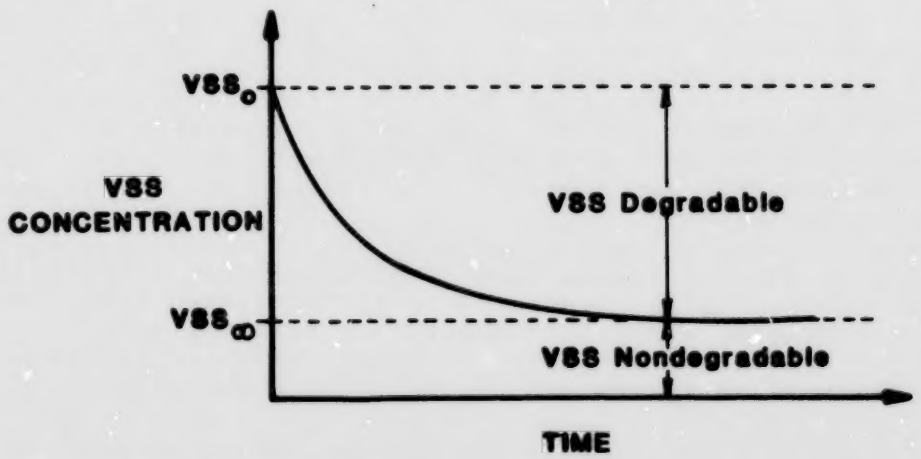

Figure 1. Batch endogenation of mixed liquor.

Degradable and nondegradable fractions are calculated based on the results of the endogenous respiration study as follows:

$$X_d = \frac{VSS_o - VSS_\infty}{VSS_o} \qquad (1)$$

$$X_n = \frac{VSS_\infty}{VSS_o} \qquad (2)$$

where; VSSo - initial volatile suspended solids concentration, mg/L.
VSS_∞ - terminal volatile suspended solids concentration, mg/L.
X_d - degradable fraction, dimensionless
X_n - nondegradable fraction, dimensionless

Studies have shown that for practical purposes the degradable fraction may be determined experimentally in approximately 30 days[1]. The degradable fraction has been found to vary from 0.8 to 0.40 for various wastes and sludge ages ranging from 2 to 40 days. The importance of degradable fraction in excess sludge generation has been incorporated in a design model by Eckenfelder (2) and will be illustrated in this paper. Degradable fraction as an activity parameter with also be illustrated in a kinetic expression for the activated sludge process.

The Degradable and Active Fraction Concepts

Degradable fraction has been proposed as an indicator of the viable proportion of a systems' biomass in soluble substrates by Upadhyaya (3). Upadhyaya demonstrated good correlation between the degradable fraction, ATP concentration, oxygen uptake rates and dehydrogenase enzyme concentration versus sludge ages ranging from 10 to 40 days in steady state continuous reactors. If we accept that degradable fraction is an indicator of the active or viable biomass in a system as demonstrated by Upadhyoya then by inference the active fraction of a system's biomass will vary with sludge age. This effect has been incorporated into the kinetic design equations for biological systems.

The degradable fraction is of importance when determining excess sludge quantities generated by the activated sludge process. However for kinetic evaluation the active fraction will be used in kinetic correlations. The active fraction is the fraction of total viable cell mass in system and can be defined as follows:

$$X_a = \frac{X_d}{X_d^i} \qquad (3)$$

where; X_a - is the active fraction, dimensionless
X_d^a - is the degradable fraction upon generation, dimensionless.

[1] The degradable fraction relates to the time frame of the activated sludge process.

X_d' is the degradable portion of a newly generated biomass and is theorectically defined as the degradable fraction of the biomass at a sludge age of zero days. The determination and typical range of values of X_d' will be provided later in this paper.

Mathematical Development

A relatively simple relationship between degradable fraction and sludge age can be developed by mass balance techniques. Equations 4 and 5 are obtained from degradable and nondegradable mass balances around a completely mixed biological reactor with a soluble substrate.

$$X_d \Delta X_v = aX_d' S_r - bX_d X_v t \quad (4)$$

$$X_n \Delta X_v = aX_n' S_r \quad (5)$$

where; ΔX_v - net change in mixed liquor volatile suspended solids, mg/L
S_r - substrate removal across reactor, mg/L
a - cell yield coefficient mgVSS/mgBOD
b - endogenous decay rate, days^{-1}
X_n' - nondegradable fraction upon generation, dimensionless

Solving for S_r in Equation 5 and substituting into Equation 4 and rearranging we obtain Equation 6.

$$\frac{1}{\theta_c} = \frac{X_d' X_n}{X_n' X_d \theta_c} - b \quad (6)$$

Using the identity, $X_d + S_n = 1$, we arrive at the following form of Equation 7.

$$1 = \frac{X_d'(1 - X_d)}{X_n' X_d} - b\theta_c \quad (7)$$

With further rearrangement and use of the identity, $X_d' + X_n' = 1$, we obtain the final form shown in Equation 8.

$$X_d = \frac{X_d'}{1 + bX_n' \theta_c} \quad (8)$$

By inverting Equation (8) a linear form between $1/X_d$ and θc may be obtained

$$\frac{1}{X_d} = \frac{bX'_n}{X'_d}\ \theta_c + \frac{1}{X'_d} \qquad (9)$$

Figure 2 shows the correlation between degradable fraction and sludge age using data generated by Eckhoff and Jenkins (4).

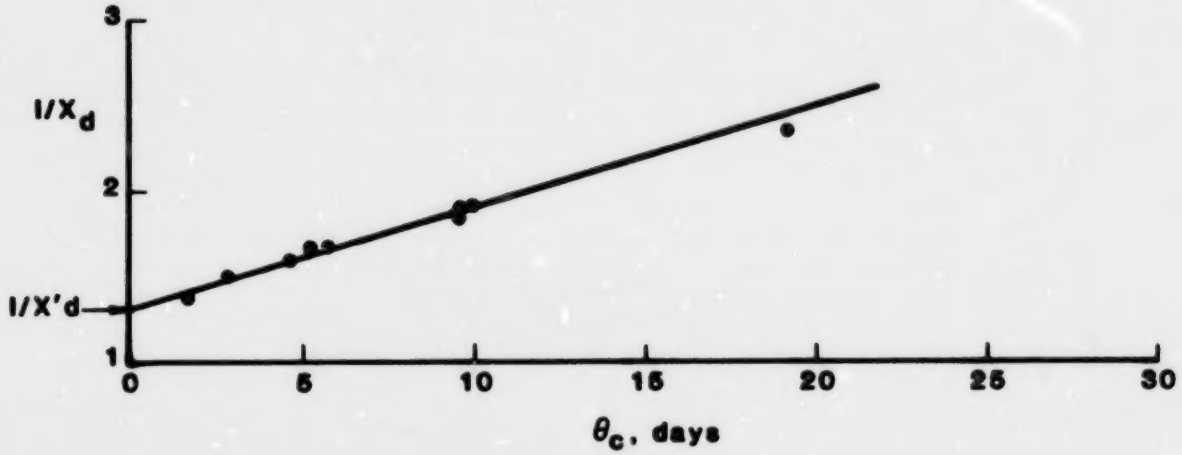

Figure 2. Linear relation between degradable fraction and sludge age.

The relationship between the degradable fraction and sludge are illustrated in Figure 3 using typical values of the endogenous decay rate and degradable fraction upon generation.

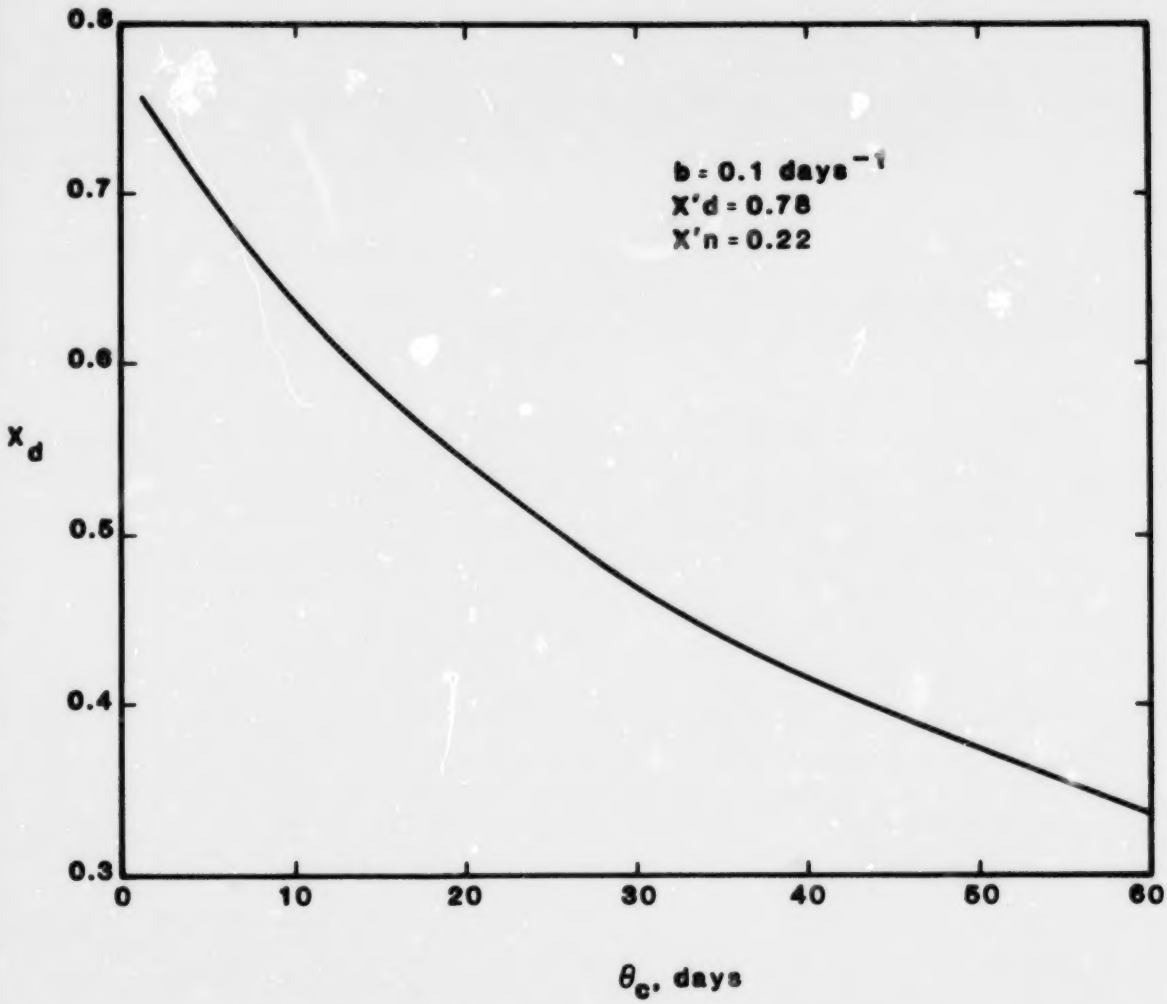

Figure 3. Degradable Fraction versus Sludge Age

Several values of X_d' are presented in Table 1 determined from data of various investigations using soluble substrates. X_d' is obtained from the y-intercept of the linear regression of Equation (9).

Table 1

Experimentally Derived Values of X_d' and X_n'

Substrate	X_d'	X_n'	Data Referenced
Skim Milk and Phenol	0.778	0.222	(5)
Glucose, phenol, and sulfanilic acid	0.750	0.250	(6)
Egg Albumen and Beef Extract	0.781	0.219	(4)
Skim Milk	0.810	0.190	(3)

An example will be provided at the end of this paper to illustrate the calculation of excess biological solids production and temperature effects.

Kinetic Model

Established kinetic models for the activated sludge process have historically failed to incorporate substrate utilization rates which are dependent on the active mass. Eckenfelders' first order kinetic model (2) will serve as an illustration of this point and is presented as Equation 10.

$$\frac{S_r}{X_v t} = K \frac{S_e}{S_o} \tag{10}$$

Here the specific substrate utilization rate, $S_r/X_v t$, is dependent on the mixed liquor volatile suspended solids concentration. Equation (4) which was derived from a mass balance of degradable solids is rewritten below as Equation (11) to illustrate the effect of active mass on growth kinetics.

$$\frac{1}{\theta_c} = a \frac{S_r}{X_a X_v t} - b \tag{11}$$

It now becomes apparent that the specific utilization rate must incorporate active mass instead of volatile mass. We can now rewrite Equation (10) based on active mass as follows:

$$\frac{S_r}{X_a X_v t} = K^1 \frac{S_e}{S_o} \tag{12}$$

It is believed that equation (12) more accurately describes

substrate utilization kinetics because removal is based on active mass and not just volatile mass. This concept is intuitively obvious upon studying Figure 2, where it is seen that the biological viability of a system is significantly decreased at higher sludge ages. An example design calculation will be provided.

Example Calculations

The following example calculations will serve to illustrate 1) excess biological solids generation and 2) activated sludge design based on the new fraction model.

Example 1 Excess Biological Solids Generation

Given the following summer and winter parameters, on a BOD_5 basis, for an activated sludge plant the excess biological solids generation will be calculated.

	Summer	Winter
K^1, $days^{-1}$	5	--
b, $days^{-1}$	0.12	--
X'_d	0.8	0.8
S_o	200	200
S_e	20	20
Temperature, ^oC	24	10
θ_k	1.065	1.065
θ_b	1.04	1.04

Summer Conditions

From Equation (12) calculate S_r/X_aX_vt

$$\frac{S_r}{X_aX_vt} = K^1\frac{S_e}{S_o} = 5(\frac{20}{200}) = 0.5$$

Calculate θ_c using Equation (11)

$$\frac{1}{\theta_c} = a\ \frac{S_r}{X_aX_vt} - b = .6(.5)-0.12 = 0.18$$

$$\theta_c = 5.6 \text{ days}$$

Calculate X_d from Equation (8)

$$X_d = \frac{X'_d}{1+bX'_n\theta_c} = \frac{0.8}{1+(.12)(.2)(5.6)} = 0.71$$

$$X_a = \frac{X_d}{X_d'} = \frac{0.71}{0.80} = 0.88$$

Calcuate $X_v t$ from $S_r / X_a X_v t$

$$X_v t = \frac{S_r}{X_a (K^1 \frac{S_e}{S_o})} = \frac{200-20}{0.88(0.5)} = 408$$

Calculate ΔX_v

$$\Delta X_v = \frac{X_v t}{\theta_c} = \frac{408}{5.6} = \underline{\underline{73 \text{ mg/L}}}$$

Winter Conditions

Correct K^1 and b for temperature

$$K^1_{10\,^oC} = 5.0\ (1.065)^{10-24} = 2.07 \text{ days}^{-1}$$

$$b_{10\,^oC} = 0.12(1.04)^{10-24} = 0.069 \text{ days}^{-1}$$

Calculate $S_r / X_a X_v t$

$$\frac{S_r}{X_a X_v t} = 2.07(\frac{20}{200}) = 0.207$$

Calculate θ_c

$$\frac{1}{\theta_c} = 0.6(0.207) - 0.069 = 0.055$$

$$\theta_c = 18.1 \text{ days}$$

Calculate X_a

$$X_d = \frac{0.8}{1+(0.069)(.2)(18.1)} = 0.64$$

$$X_a = \frac{0.64}{0.80} = 0.80$$

Calculate $X_v t$

$$X_v t = \frac{180}{0.80(0.207)} = 1{,}087$$

Calculate ΔX_v

$$\Delta X_v = \frac{1087}{18.1} = \underline{\underline{60 \text{ mg/L}}}$$

Example 2 Active Fraction Kinetic Model

Given that the following kinetic data has been determined from

laboratory studies calculate the required hydraulic retention time for a steady state treatment efficiency.

Given; $K^1 = 5.2 \text{ days}^{-1}$
$a = 0.6$
$b = 0.1$
$S_o = 300$
$S_o^e = 20$
$X_d^f = 0.8$

Calculate the quantity $S_r/X_a X_v t$ using Equation (12)

$$\frac{S_r}{X_a X_v t} = K^1 \frac{S_e}{S_o} = 5.2\left(\frac{20}{300}\right) = 0.35$$

Calculate sludge age from Equation (11)

$$\frac{1}{\theta_c} = a \frac{S_r}{X_a X_v t} - b = (0.6)(0.35) - 0.1 = 0.108$$

$$\theta_c = 9.3 \text{ days}$$

Calculate X_d from Equation (8)

$$X_d = \frac{X_d'}{1+bX_n' \theta_c} = \frac{.8}{1+(0.1)(0.2)(9.3)} = 0.68$$

$$X_a = \frac{X_d}{X_d'} = 0.84$$

Calculate $X_v t$ from value of $S_r/X_a X_v t$

$$X_v t = \frac{S_r}{X_a\left(K^1 \frac{S_e}{S_o}\right)} = \frac{300-20}{0.84(0.35)} = 952 \frac{\text{mg-day}}{\text{L}}$$

Assume a reasonable operational value of X_v, say 2000 mg/L and calculate t

$$t = \frac{952}{2000} = \underline{\underline{0.48 \text{ days}}}$$

It should be noted that when temperature effects account for changes in the kinetic rate constants both K and b must be adjusted accordingly. It may be shown that the effect of the active mass model is not as significant when the rate constant, K, changes substantially with temperature due to the compensating endogenous decay rate change with temperature.

DISCUSSION

A simplified correlation between degradable fraction and sludge age is presented. This relationship is in a more useable form than previous correlations between degradable fraction and food to microorganism ratio developed by Eckenfelder (12). Use of the degradable fraction is critical in obtaining accurate estimates of excess biological solids generation.

A modified kinetic formulation based on active fraction is also presented and has been illustrated to produce significantly different results than the conventional model in moderate to high sludge age completely mixed biological systems at steady state.

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DEWATERING OF POLYMER CONDITIONED MUNICIPAL SLUDGES USING MEMBRANE FILTER PRESSES

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G.G. Boudaud, C.E.T.

INTRODUCTION

Duffin Creek Pollution Control Centre is the treatment component for the 132 Kilometre-long York-Durham Sewerage System which will ultimately be one of the largest collection and treatment systems in Canada. The system serves several political jurisdictions and is designed to have an ultimate average daily flow of 842,000 m^3/d and serve a population close to 900,000 plus some 6,880 hectares of industrial land.

The Duffin Creek Pollution Control Centre was put into operation in April 1979 with a sludge incineration capability commencing in 1981. The fluidized-bed incineration system is equipped with combustion air pre-heaters and waste heat recovery boilers.

In 1984, work began in the installation of four membrane sludge filter presses for sludge dewatering at Duffin Creek. They will be the first of their kind in North America dewatering polymer conditioned sludge. But, although new to this continent, these systems have been employed extensively in England and Europe. The design work at Duffin Creek will bring together several engineering components which have proved highly reliable in operation while being economical to operate.

While the purchase and installation of the four presses is expected to cost some $2.8 million the energy savings are expected to recoup the capital cost in about four years.

This paper will discuss the membrane filter presses being installed at Duffin Creek and the British Experience as well as the pilot work performed at other treatment plants while the system was being evaluated.

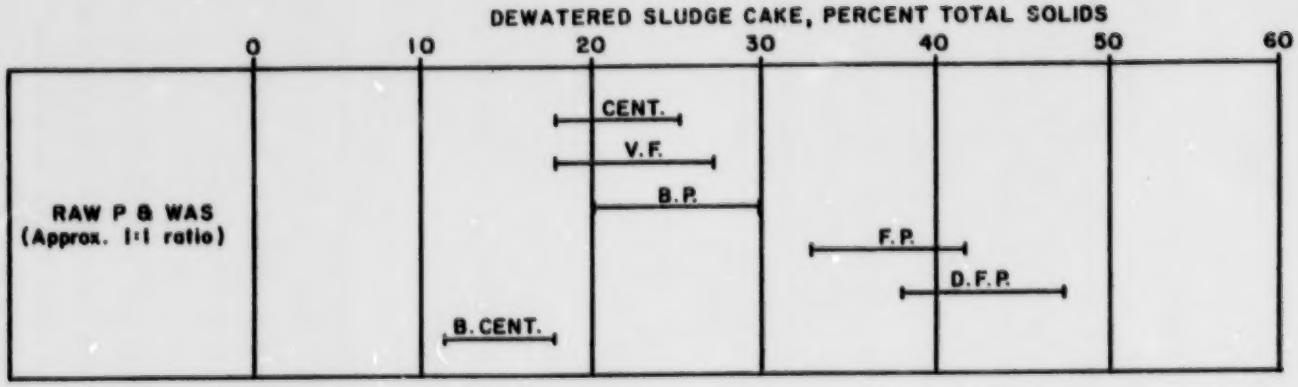
CURRENT TECHNOLOGY

Dewatering of sewage sludges presents perhaps the most difficult problem in the design and operation of a waste water treatment plant. Dewatering is normally required prior to ultimate disposal by landfill, barging to sea or incineration. Throughout the years, various methods of conditioning and dewatering have been used with varying degrees of efficiency and cost effectiveness.

Vacuum filters and plate filter presses have been used for a number of years as prime dewatering devices. These devices have proven to be durable but relatively high in capital cost. They also had the added disadvantage that the sludge normally had to be conditioned with inorganic conditioners such as ferric chloride and lime prior to dewatering. These materials created additional operating problems at the plant and did nothing to assist the incineration process.

As polymers for sludge conditioning became available, the use of centrifuges and belt filter presses became more popular. Polymer conditioning has proven to be less troublesome from an operating standpoint and cost effective when used in conjunction with belt filter presses, centrifuges and in some cases, vacuum filters.

FIGURE 1



LEGEND

CENT.	– SOLID BOWL CENTRIFUGE	F.P.	– FILTER PRESS
V.F.	– VACUUM FILTER	D.F.P.	– DIAPHRAGM FILTER PRESS
B.P.	– BELT PRESS	P	– PRIMARY
B. CENT.	– BASKET CENTRIFUGE	WAS	– WASTE ACTIVATED SLUDGE

DEWATERED SLUDGE CAKE PERCENT SOLIDS FOR MIXTURES OF RAW PRIMARY AND SECONDARY SLUDGES

(USEPA - SLUDGE TREATMENT AND DISPOSAL)

At this point it may be useful to look at two figures. The first is extracted from the US EPA Design Manual on Dewatering Municipal Wastewater Sludges. You will note that centrifuges, vacuum filters and belt presses all are anticipated to produce cake solids in the range of 17 - 30% for mixtures of raw primary plus waste activated sludge. On the other hand filter presses and diaphragm filter presses are anticipated to produce sludge cakes in the range of 33 to 48% on similar sludge mixtures.

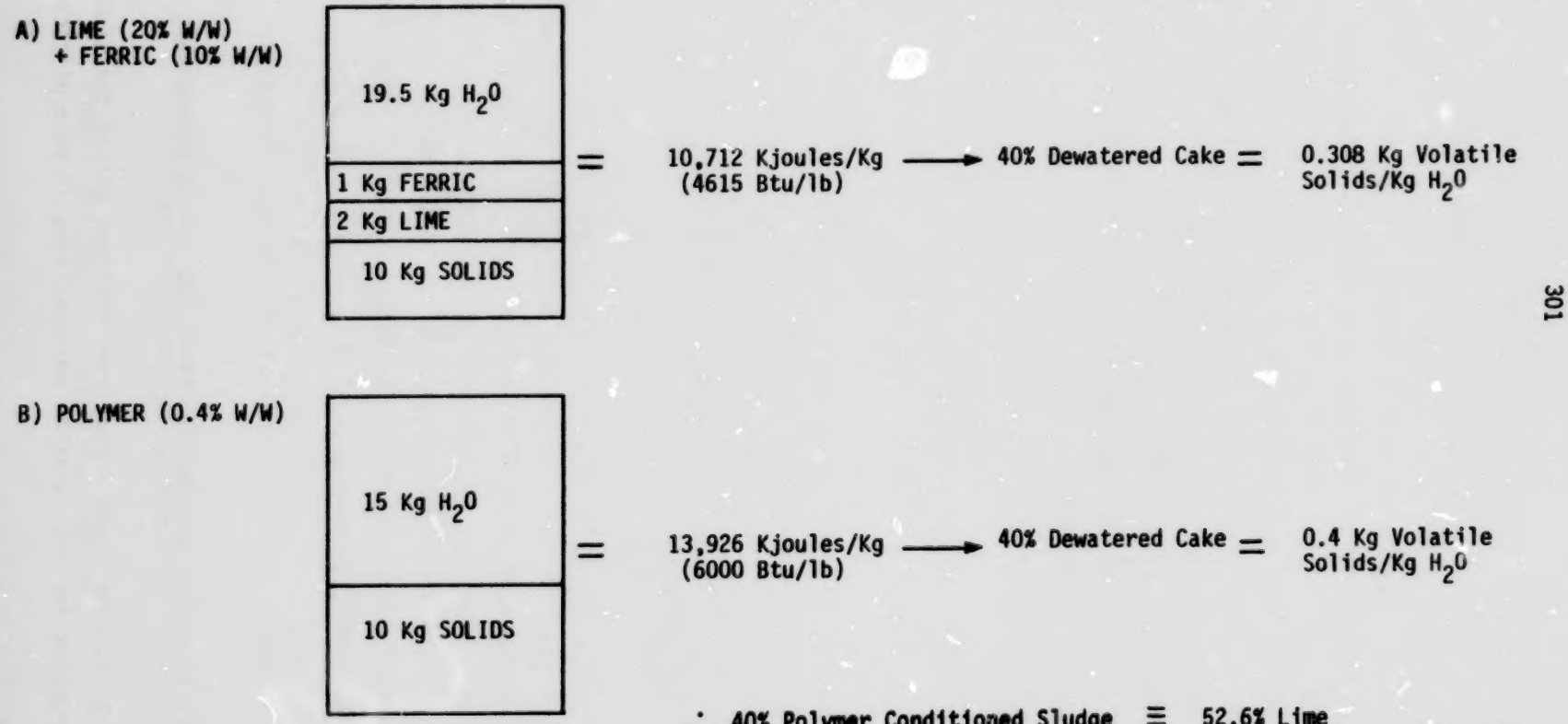
If one is planning to landfill or incinerate the sludge then the significantly drier sludge cake has tremendous advantages. However, one must be certain that the additional cake dryness is not simply a result of the addition of inert material. Figure 2 shows the implication of ferric chloride and lime conditioning versus polymer conditioning on a 40% sludge cake.

EXPERIENCE AT THE DUFFIN CREEK PLANT

In the mid-seventies the Ministry of the Environment of Ontario became involved in the review of sludge dewatering technology for use in plants utilizing incinerators as the method of ultimate disposal. Serious consideration was given to filter presses. However, bench scale pilot tests utilizing polymer as a conditioner did not show any significant promise. The decision was subsequently made to install belt filter presses at the Duffin Creek Water Pollution Control Plant.

Three years of incineration operation at the Duffin Creek Plant proved that it is not possible to incinerate the belt filter press cake without the addition of auxiliary fuel. The presses could obtain cake solids of between 17 and 26 percent but the older style presses experienced high maintenance problems when operated to produce the higher cake solids.

FIGURE 2


DEWATERED SLUDGE CAKE
(40% D.S. + 60% VOLATILE SOLIDS)
A) LIME (20% W/W) + FERRIC (10% W/W)
19.5 Kg H_2O
1 Kg FERRIC
2 Kg LIME
10 Kg SOLIDS
=
10,712 Kjoules/Kg (4615 Btu/lb)
40% Dewatered Cake
=
0.308 Kg Volatile Solids/Kg H_2O
B) POLYMER (0.4% W/W)
15 Kg H_2O
10 Kg SOLIDS
=
13,926 Kjoules/Kg (6000 Btu/lb)
40% Dewatered Cake
=
0.4 Kg Volatile Solids/Kg H_2O
∴ 40% Polymer Conditioned Sludge ≡ 52.6% Lime + Ferric Conditioned Sludge

Additional dewatering units were required at the Duffin Creek Plant. At the same time, reports persisted from the U.K. that polymer conditioning was being successfully used in conjunction with filter presses to produce a dry cake with no operating difficulties. The decision was made to investigate the use of filter presses in conjunction with polymer conditioned sludge.

A bench scale pilot test was performed at Duffin Creek utilizing a small pilot unit. These results indicated that sludge cake in excess of 40% was attainable.

In view of the enormous cost saving potential it was further decided to conduct a review of operations of the U.K. installations.

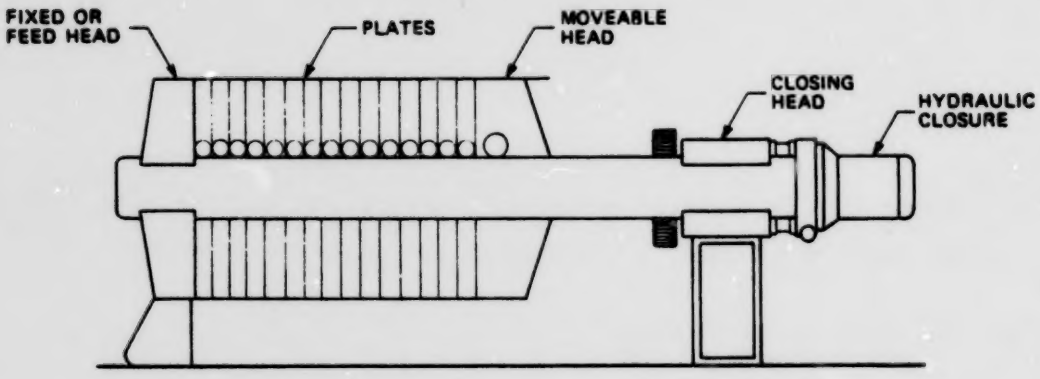
Upon completion of the review it was evident that there had been a significant evolution in the use of filter presses with polymer conditioned sludges.

THE BRITISH EXPERIENCE

Pressure filtration has probably received more critical examination than any other single type of dewatering device. Its ability to produce a very dry end product compared with other systems is the prime reason for continued interest in this device. A schematic diagram of a traditional filter press has been extracted from the US EPA Manual on Sludge Treatment and Disposal and is presented as Figures 4 and 5. These units have certain inherent disadvantages:-

- the system is a batch operation with limited automation;
- cycle times can be lengthy, resulting in low throughput;
- inorganic conditioners have been employed making the sludge cake less attractive for incineration and resulting in more difficult chemical

FIGURE 3



SCHEMATIC SIDE VIEW OF A RECESSED PLATE PRESSURE FILTER

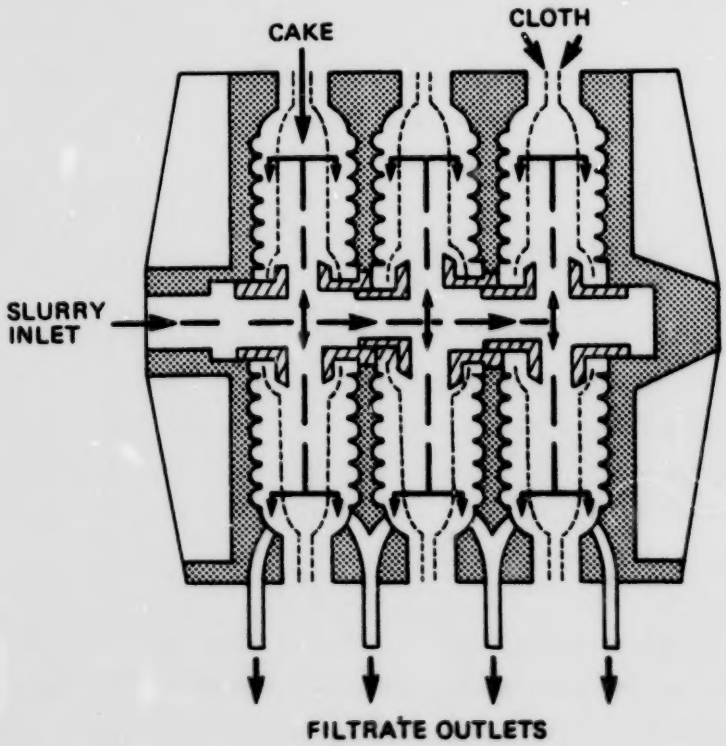


FIGURE 4

CROSS SECTION OF A FIXED-VOLUME
RECESSED PLATE FILTER ASSEMBLY

(USEPA - SLUDGE TREATMENT AND DISPOSAL)

handling.

In recognition of these disadvantages, several changes have occurred in recent years which tend to deminish or eliminate the problems.

(1) Automation of Presses

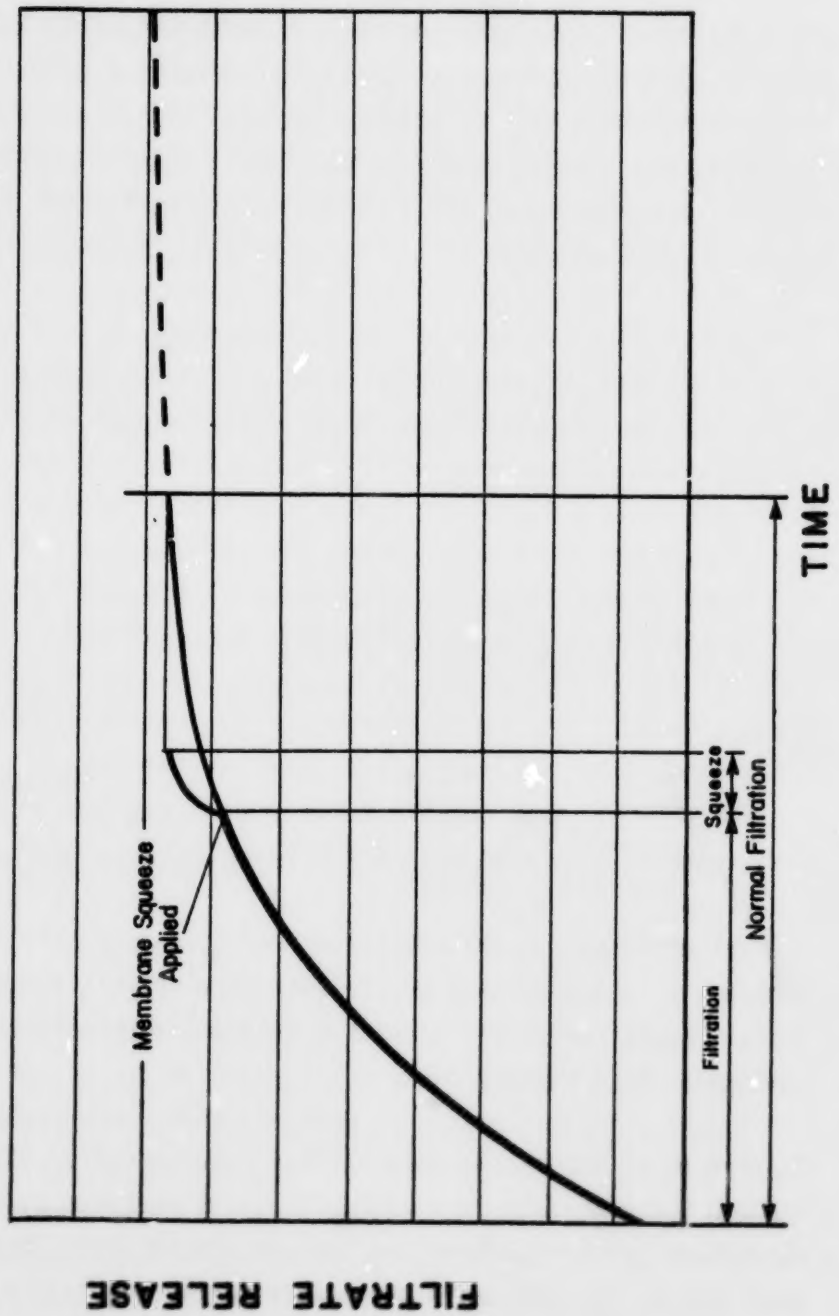
Manufacturers of filter presses recognized the necessity to redesign and modernize their product to reduce the reliance on manual labour. Modern presses now provide automatic plate shifting devices, automatic plate washers and programmable controllers to minimize operating input. These advances significantly reduce operating and maintenance on the presses. However, the process remains a batch operation - a disadvantage that can be greatly deminished by reduction of the cycle time.

(2) Reduction of Cycle Time

The pressure filtration phase in a normal press cycle is the largest time consumer in the sequence of operations. In the traditional fixed volume press, times in the range of 4 - 6 hours are relatively common using inorganic conditioners. Polymer conditioned sludge, however, releases 70 - 80% of its filtrate in the first 30 or 40 minutes of pressing.

To terminate the filtration cycle after only 80% of the filtrate has been added would double the performance efficiency. Using conventional recess plates for pressing, this is not possible as a completely unacceptable cake would result. However, using the membrane plate it is possible to terminate the filtration cycle at this point. The remaining 20% can be removed during the membrane squeeze cycle. Figure 5 shows a typical time - filtrate release curve for a polymer conditioned sludge.

FIGURE 5



TYPICAL RATE OF RELEASE CURVE

The membrane plate consists of a rubber membrane moulded around a flat steel sheet insert, the rubber membrane forms a complete rubber envelope around the steel insert. The membrane face of the plate is moulded with a drainage surface such as dimples or channels. The steel insert is provided with an air inlet so that compressed air can be applied to the rubber envelope, thereby inflating both faces of the membrane plate. The membrane plate is constructed with drainage facilities and can have provision for centre or corner feed.

Drainage ports that are provided in each plate collect the filtrate and convey it to the end of the press where it is discharged into a common drain. The processing cycle commences with conditioned sludge being fed at a rapid rate until all chambers are filled. This rate falls off as cake begins to form. Pressure filtration then takes place at the selected pressure i.e. between 500 - 850 kPa, until sufficient sludge is fed to the press at which point air is introduced into a membrane within each alternate plate.

The plate was originally intended for application in the chemical preparation industry and has been used successfully in this context for many years. However, the plate was not applied to the dewatering of sewage sludges until 1973.

A press would be equipped with a series of plates - recess plates and membrane plates alternating - so that each chamber has one fixed face. The following describes the method of operation of a press equipped with these plates.

At some point during the normal filtration operation the feed pump is stopped and the membrane surfaces are then inflated by applying compressed air pressure at the back of the membrane faces. The wet cake within the chamber is then squeezed so that liquid within the chamber is forced under pressure through the filter cloths. The membrane surface will take up a shape determined by the amount of

sludge solids held within the chamber.

The squeeze can be maintained at a pressure not exceeding 1,000 kPa for a period sufficient to produce an acceptable cake. The air pressure may exceed even the maximum of 1,000 kPa used in these tests. The dryness of the cake is determined by the duration of the squeeze and the pressure applied. At the end of the predetermined period of squeeze, the air pressure is exhausted and the sludge cake removed from the press.

The pressures are later increased to the normal recess press feed pressure of 700 kPa and an applied air squeeze pressure of 1,000 kPa. Clearly by using this system a filtration cycle can be terminated after the addition of 75 percent of the solids desired to produce a cake which would be satisfactory under a normal recess system of pressing. The cake produced will be thinner but can have a higher DS content. Thus the membrane plate system offers the means of operating presses using a short duration pressing cycle.

(3) Sludge Conditioning

(a) Polymer Development

The new series of high molecular weight, cationic polyelectrolytes have been used with significant benefits as sewage sludge dewatering aids on the filter presses. Originally, ferric chloride and lime were used as sludge conditioners in the design of a filter press system.

The major part of the research and development of utilizing polymer as a conditioning agent for the filter press system has been carried out by Allied Colloids Ltd. of the U.K. This company has developed polymers with varying degrees of cationic activity and having molecular weights varying from (5.0 - 10) x

10^6, specifically for this type of sludge dewatering process. Process advantages that have been gained by utilizing polymer have included lower filtration cycle times, lower treatment costs, lower ash values in the cake, reduction in size of equipment required and lower labour and maintenance costs.

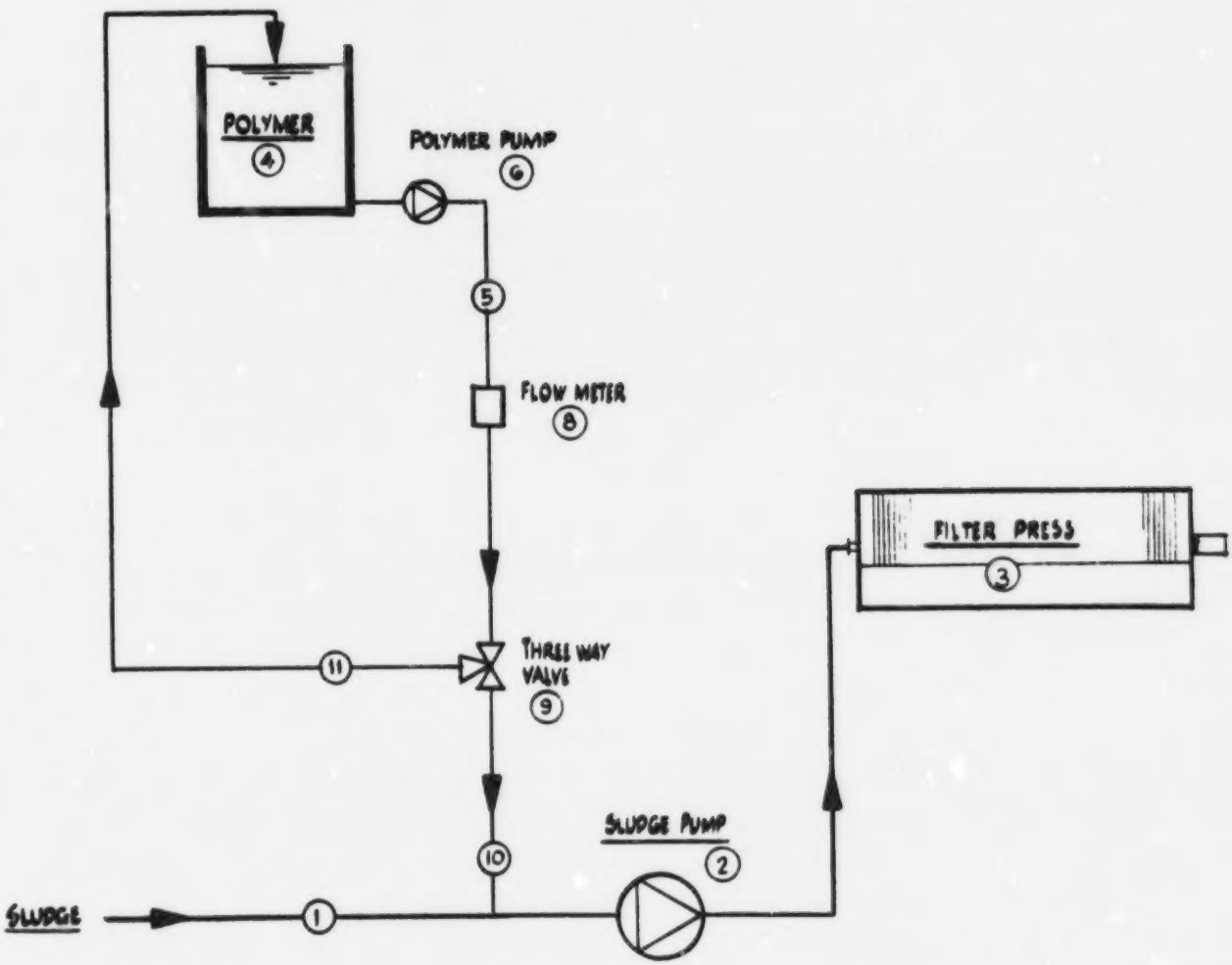
The selection of a particular polymer for any sludge depends on many factors. Although it is possible to generally predict performance it is advised that a series of selection tests be performed on the sludge in question, including a simple capillary suction time (C.S.T.). evaluation and small pilot scale pressure filtration tests.

The successful use of a polymer for the pressure filtration of sludge can be as dependent on the method of application as on the product itself. Inadequacies of the addition system have accounted for many past failures in sludge pressing with polymer. This is especially true where the chemical is applied on a batch basis to a volume of sludge contained within a conditioning tank. In many cases, a combination of inadequate mixing, long sludge retention times in the conditioning tank and floc breakdown within the tank have produced inferior results. These inadequacies, however, have resulted in the development of a novel "in-line" procedure which has eliminated most of the earlier drawbacks of the process.

(b) <u>In-line Polyelectrolyte Addition</u>

The "in-line" polymer dosing system was introduced by Allied Colloids in 1974 in the U.K.

FIGURE 6



POLYMETER SYSTEM

The dosing unit named the "Polymeter" system injects a measured quantity of polymer solution into the suction side of the "Willet" ram pumps used to charge the press. Each "Polymeter" unit consists of a flowmeter and valve to preset the flow while returning the polymer to the holding tank. Limit switches set on the Willet pumps activate a solenoid valve at the beginning of the suction stroke which enables polymer to be injected into the sludge line at the preset pressure and flow. Polymer is re-directed back to the return at the end of the stroke by another limit switch. This system is shown schematically in Figure 6.

The sewage sludge is drawn through pipe [1] to an in-line Willet pump [2] by which it is forced into a filter press [3]. A supply of polymer is provided in tank [4] and this is pumped at a constant rate through pipe [5]. The constant rate is provided by conventional means, utilizing a variable speed Moyno pump [6]. The constant rate can be observed by flowmeter [8]. A three-way valve [9] is provided in pipe [5] and is electrically connected to two micro-switches fitted to pump [2]. At the beginning of the suction stroke, the three-way valve is energized in such a way as to allow flow through the pipe [10] and prevent flow through pipe [11]. At the end of the suction stroke valve [9] is de-energized to prevent flow along pipe [10] and to cause flow to be diverted along pipe [11] back to tank [4].

In this way the polymer is pumped into the sludge line only during the suction stroke of the Willet pump. By ensuring a constant flow of polymer into a constant flow of sludge, very uniform metering of the polymer into the sludge can be achieved. Since the suction stroke has a constant period, the volume of polymer added during a suction stroke is constant.

The Willet pump is positioned in the conventional in-line manner and the polymer is added to the sludge immediately prior to the Willet pump. Reliance on the suction stroke of the pump to feed polymer into the sludge gives the added advantage of controlled addition to the sludge.

Accurate metering of polymer into the sludge is provided at all times regardless of the fact that in a variable pressure system such as filter pressing, the output of the Willet pump will vary considerably. This is provided for by metering the polymer during the suction stroke, which is constant, not the delivery stroke.

PILOT STUDIES

Following preliminary cost evaluations, arrangements were made to conduct pilot studies at three treatment plants:

- Duffin Creek Pollution Control Centre - Trials to evaluate the filterability of raw co-settled primary plus waste activated sludge;

- Sault Ste. Marie East End Sewage Treatment Plant - Trials to determine the dewaterability of raw primary sludge;

- Highland Creek Sewage Treatment Plant - Trials to determine the dewaterability of heat treated sludge and of thickened waste activated sludge.

The pilot test unit consisted of two 500 mm membrane filter plates which were alternated with three 500 mm recessed plates resulting in four dewatering chambers.

a) Duffin Creek Trial

The trial was undertaken to evaluate the length of cycle time that might be expected to dewater the Duffin Creek sludge. Unfortunately, during the trial period, sludge quality at the plant was below normal with C.S.T. values in the 500 sec. range, as opposed to normal operating levels of the fresh raw primary plus waste activated sludge of 250 - 300 sec. Various cake thicknesses were investigated as well as different types of cloth material. Test data to date indicates that a 20 mm final cake thickness is the most desirable, achieving better cake release proeprties. The cloth type best suited to the Duffin Creek sludge was the nylon monofilament Rilsan type cloth, which showed good cake release properties.

The cakes produced were well formed and when left outdoors for a period of 24 hrs. during a period of intermittent rain did not deteriorate noticeably. The ability of this type of dewatered cake to shed water in this manner is due to the destruction of the capillary lines within the sludge which takes place during the pressure filtration cycle. This makes dewatered cake more easily manageable on a landfill application.

Polymer utilized for the pilot test was Percol 757 supplied by Allied Colloids Ltd. Canada. Polymer consumption was in the range 4 - 5 kg/tonne D.S. These high dosage rates were a direct result of the poor quality sludge being dewatered. Previous test data on fresh sludge indicated polymer dosages of between 3 - 4 kg/tonne D.S. (Table 1).

DUFFIN CREEK P.C.P.

TABLE 1

Run No.	Raw Sludge % D.S.	C.S.T.	Polymer Dosage kg/tonne	Dewatered Cake % D.S.	Cycle Time
1	5.75	475	4.7	31	45 min. fill 7 min. squeeze
2	5.75	420	4.3	34	30 min. fill 13 min. squeeze
3	4.9	450	4.8	30	40 min. fill 15 min. squeeze
4	5.6	485	4.8	31	25 min. fill 20 min. squeeze
5	5.8	390	4.0	39	50 min. fill 15 min. squeeze
6	5.7	440	4.5	33	30 min. fill 20 min. squeeze
7	5.5	500	5.0	30	25 min. fill 20 min. squeeze
8	5.5	470	4.6	35	30 min. fill 20 min. squeeze
9	6.1	465	4.7	33	30 min. fill 30 min. squeeze

(b) Sault Ste. Marie Trial

The trial results are shown in Table 2. These results indicate that the raw primary sludge (5%) can be dewatered to approximately 50 D.S. in about 40 minutes total fill and squeeze time. Polymer requirements for the trials were 0 - 3.9 Kg/tonne.

The resulting cake exhibited strong structural qualities. This factor was important since the sludge cake at Sault Ste. Marie is sent to a landfill site.

SAULT STE. MARIE - EAST END

TABLE 2

Test Data

Run No.	Raw Sludge % D.S.	Raw Sludge C.S.T. Sec.	Polymer Dosage kg/tonne	Dewatered Cake % D.S.	Cycle Time Fill/ Squeeze
1	6.4	70	Nil	54	1.5 hrs/1 hr.
2	6.4	30	0.55	45	30 mins/30 mins
3	6.4	50	3.4	53	15 mins/20 mins
4	5.0	65	3.8	47	20 mins/15 mins
5	5.0	60	2.7	50	25 mins/20 mins
6	5.0	75	3.9	49	20 mins/17 mins
7	5.5	70	3.6	57	20 mins/20 mins
8	4.5	65	2.8	50	20 mins/20 mins
9	4.5	60	3.2	52	12 mins/20 mins

(c) Highland Creek Trial

The Highland Creek trial was conducted to provide information on the dewatering of heat treated and thickened waste activated sludge. No project was contemplated at this plant and only four runs were performed. The test results are shown in Table 3.

The heat treated sludge was dewatered to approximately 65% D.S. in forty-five minutes with no polymer addition. The thickened waste activated sludge was dewatered to approximately 25% in one hour with the addition of 4.5 - 5 Kg/tonne polymer.

HIGHLAND CREEK S.T.P.

TABLE 3

Type of Sludge	% D.S.	Polymer Dosage	Dewatered Cake % D.S.	Cycle Time Fill/Squeeze
Heat treated sludge	10%	Nil	64%	30 mins/15 mins
Heat treated sludge	9.5%	Nil	65%	25 mins/20 mins
Thickened Waste Activated sludge	5%	5 kg/tonne	25%	25 mins/30 mins
Thickened Waste Activated sludge	6%	4.5 kg/tonne	24.5%	20 mins/40 mins

RESULTING ACTION

Based on the field observations in the U.K. and the pilot test work performed in Ontario the decision was made to install Membrane Filter Presses at the Duffin Creek Pollution Control Plant. Four presses were selected to serve the 180,000 m^3/day plant. The presses are anticipated to be in operation by mid-December 1984.

ACKNOWLEDGEMENTS

The authors wish to acknowledge the assistance and co-operation afforded to them by the following persons during the tour of the U.K. installations and the conduct of Pilot Tests in Ontario:

Tour of U.K. Installations

Ministry of the Environment (Ontario)	-	Mr. C.J.K. Wilson
Allied Colloids U.K.	-	Mr. G. Dawson
Allied Colloids Canada	-	Mr. J. Macdonald
Edwards and Jones	-	Mr. M. Hulme
Asdor Ltd.	-	Mr. A.S. Doran

Pilot Tests

Regional Municipality of Durham	-	Mr. A. Leitch
(Duffin Creek)	-	Mr. J. Clarke
	-	Mr. B. Kent
City of Sault Ste. Marie	-	Mr. D. Redmond
	-	Mr. R. Weldon
Municipality of Metropolitan Toronto	-	Mr. D. Clough
	-	Mr. P. Wimmer

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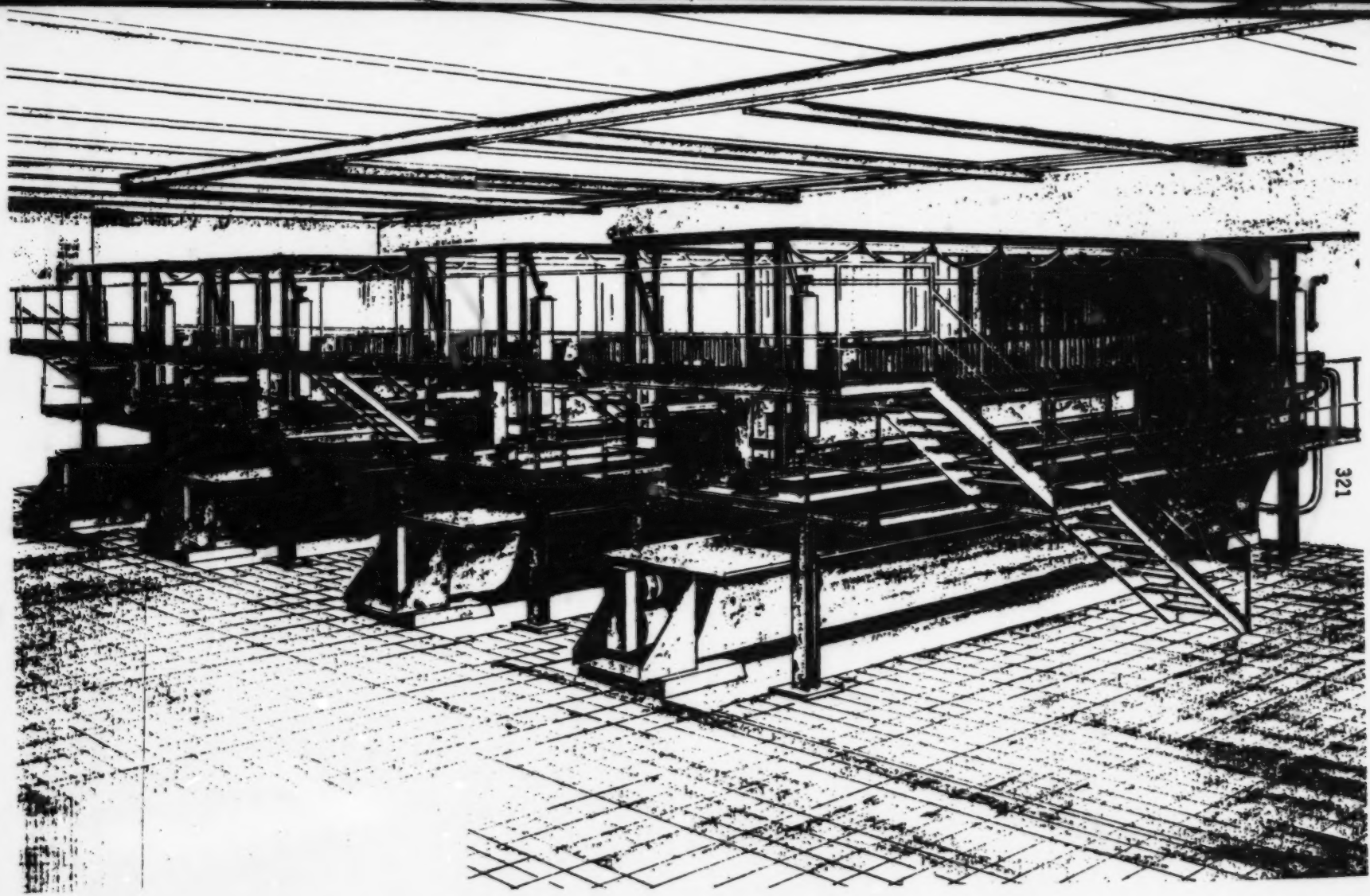

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ARTIST'S RENDITION OF FILTER PRESS OPERATION

COST BENEFITS GENERATED BY REPLACEMENT OF VACUUM FILTERS WITH FILTER BELT PRESSES

WILL HAAPALA
OPERATIONS SUPERINTENDENT
WESTERN LAKE SUPERIOR SANITARY DISTRICT

AND
TERRY ZAUDTKE, P.E.
CONKLIN, PORTER AND HOLMES, ENGINEERS, INC.

A. INTRODUCTION

The 1.91 m^3/s MGD design flow Western Lake Superior Sanitary District wastewater treatment facility serves industries and municipalities in a 1295 square kilometer area of Northeastern Minnesota. Plant processes consist of preliminary treatment, Unox pure oxygen activated sludge, flocculation, gravity filtration, chlorination and dechlorination. The facility has consistently produced effluent quality well below NPDES permit limits and received O&M awards from U.S.E.P.A. for 1981 and M.P.C.A. for 1982.

Waste activated sludge produced by the Unox process (27-32 metric tons per day) is flotation thickened and belt press dewatered in preparation for land application or fluidized bed incineration disposal processes, which are alternated on a seasonal basis. A full-scale land application program was developed to provide sludge disposal until the incinerators became routinely operational in November 1982. At the present time, land application from May 1 - November 1 and incineration the remainder of the year provide the most economical mode of operation for the District facility.

Original design of the WLSSD plant, completed in 1974, included the conventional vacuum filter process. Four Komline-Sanderson Flexibelt filters were provided for dewatering, using inorganic conditioning. Plant construction was concurrent with growing acceptance and development of the continuous belt filter press, however, the belt press option was not considered viable during the 1974 design period.

Startup of the new WLSSD facility occurred in September 1978. After a six-month shakedown, the vacuum filtration system had stabilized to a point at which solids handling allowed effective operation of the wastewater treatment process. However, several difficult problems were inherent to the vacuum filter process, including blockage of process and drain lines with lime and iron scale, safety hazards relating to the operation, wear and failure of lime systems, and corrosion of piping and tanks. These problems and relatively high energy, chemical and labor costs associated with filtration justified a change in dewatering process.

WLSSD initiated a project in mid 1980 to investigate and test alternate dewatering processes.

B. Pilot Studies

Extensive preliminary work, including literature review, supplier contacts, visits, and calls to other facilities, was carried out by WLSSD staff. As a result, the continuous filter belt press was selected as a probable replacement for vacuum filtration.

The belt press process was attractive in that it offered lower O&M costs than the existing system and was predicted to use only polymer conditioning, which would produce a high volatile content, low ash sludge cake for incineration. To test feasibility of the filter belt press process, WLSSD carried out three pilot tests. Participating suppliers were Ashbrook-Simon-Hartley (Klampress, 1.0 m, October 1980), Arus-Andritz (CPF - S7 1.0 m, April 1981) and Filtration Technology (FT-50" October 1981).

All three tests proved that the filter belt press process would provide improved dewatering operation and a daily O&M cost saving of $300-$500 over vacuum filtration. The predicted rapid payback of this process change led to the creation of a fairly innovative engineering project for implementation of the process change. Tables #1 and #2 list results generated from the pilot tests.

Table #1

PERFORMANCE COMPARISON, VACUUM
FILTER TO DEWATERING PRESS
OCTOBER 1980 TEST

	VACUUM FILTER (1)	PRESS (2) (1.5 m. width)
Metric tons per day	20.1	18.0
Per Unit D.S.	18.2	16.3
Filter Cake		
%TS	18.9%	18.1%
%TVS	60.4%	83.4%
Fuel value	3392	5338 Gm.cal./gm.
% Recovery	93%	90%
Connected H.P.	101.4	15.21
Cake Production, metric Tons/Day	129	90
Ash Quantity, metric Tons/Day	9.6	2.7

(1) 1980 Experience

(2) October 1980 Test Results

PERFORMANCE COMPARISON
VACUUM FILTER TO BELT PRESS
APRIL 1981 TEST

	VACUUM FILTER (1)	PRESS (2) (54" Effective Width)
Metric Tons D.S. per day per unit	18.2	18.1
D.S.		
Filter Cake		
%TS	18.9%	18.3%
%TVS	60.4%	79.3%
Fuel value	3392 Gm. cal/gm.	4765 Gm.cal./gm.
% Recovery	93%	92.7%
Connected metric H.P.	101.4	15.21
Cake Production, metric Tons/Day	129	99
Ash Quantity, metric Tons/Day	9.6	3.6

(1) 1980 WLSSD Experience

(2) April 1981 test results, adjusted to 18.1 metric tons/day D.S. throughput

WLSSD FILTEC 1.8 M. FILTER BELT PRESS TRIAL

OCTOBER 7-9, 1981

TABLE #2

DATE	RUN NO.	FEED		SLUDGE	POLYMER				FILTER CAKE		FILTRATE-BELT WASH			%RECOVERY
		L/sec	%TS	Kg./hr.D.S.	Type	Conc.	Kg./ton	$/ton	%TS	%TVS	Flow L/sec	T.S. mg/l	TSS/mg/l	
10/8	1	4.4	7.15	1136	Hercofloc 812	0.5%	6.3	$24.17	17.6	69.0	6.6	1070	288	97.7%
	2	4.4	6.78	1077	812	0.5%	7.1	$27.43	17.5	76.3	6.7	1185	382	97.1%
10/9	3	2.8	5.46	557	812	0.5%	9.5	$36.64	17.1	77.3	4.0	1178	128	96.9%
	4	2.8	7.12	727	812	0.5%	7.8	$29.92	22.2	76.9	4.0	1076	92	97.8%
	5	3.2	--	--	812	0.5%	--	--	22.5	77.5	--	--	--	--
	6	3.2	--	--	812	0.5%	--	--	16.7	77.3	--	1360	404	--

C. Project Engineering

The payback period for the belt press installation predicted by the pilot tests was estimated at one to three years. To take advantage of this cost benefit as rapidly as possible, and minimize the possibility of an unsuccessful installation, WLSSD Engineering designed the project with the following features:

1. Local funding
2. In-house design and inspection
3. Turn-key construction with Supplier as General Contractor
4. Selection of Supplier/Contractor based on bid price, performance guarantee, mechanical/structural analysis, supplier presentation, present worth analysis, contract completion time
5. Payment Schedule including 50% withholding through completion of performance tests, and $500 per day liquidated damage clause for non-performance

These provisions, again, were intended to assure rapid installation of an effective system, and selection of quality equipment.

Filtration Technology, Inc. was awarded the belt press contract on November 19, 1981, with a 104 day completion bid.

Based on the pilot studies, Filtec bid two FT-75 1.8 m operating width belt presses, with control systems and polymer storage and metering equipment. Final design of the system was a joint effort between WLSSD Engineering and Filtec. Mechanical and electrical work was sub-contracted to local firms. Installation of a Serpentix sludge conveyor to serve the presses was done by direct purchase and installation by separate contract.

Major equipment was delivered in February 1982, construction was complete by mid-April and continuous beneficial operation began on May 19, 1982. The final performance test was completed on September 1, 1982. The project was successful in providing an operating system in a relatively short period of time. Optimization of the system lasted through 1983, with several electrical, mechanical and building system changes required.

Figures #1 and #2 list the contract bids and performance guarantees, and WLSSD rating system for bid selection.

FIGURE #1

Item No.	Description	WLSSD BID NO. 199 Engineer's Estimate	BIDDER Andritz	Ashbrook	Envirex	Option 1 Filtec	Option 2 Filtec
1	Lump Sum Bid for Belt Filter Presses (dollars U.S.)	350,000	486,300	452,874	425,900	401,228	421,228
	Deduct for different payment schedule			30,608	14,975		
2	Contract Completion Time (days)	220	220	220	220	104	104
3	Number of Belt Presses to be furnished	2	2	2	2	2	2
4	Guaranteed Performance						
	A. Total Sludge Feed Rate (L/sec)	9.46	9.46	11.35	9.46	9.46	11.04
	B. Total lbs. in Sludge Feed (Kg/day)	54422	54422	58957	54422	54422	63492
	C. Total Kg in Cake (Kg/day)	51700	51700	56009	51700	51700	60317
	D. Capture (%)	95	95	95	95	95	95
	E. Polymer Dosage (Kg/ton)	8	7	8	8	8	8
	F. Cake Solids Concentration (%)	18	18	18	18	18	18

Figure #2

WLSSD Evaluation Bid No. 199

Vendor

ITEM	ASSIGNED POINTS	Filtec	Envirex	Ashbrook	Andritz
1. FUNCTIONAL DESIGN	40				
a. Belt tensioning system	6				
b. Ease of Belt Replacement	4				
c. Overall Size of Machine	6				
d. Feed Distribution System & Gravity Drainage Section	8				
e. Wedge Section	5				
f. Functional Doctor Blade	6				
g. Belt Washing System	5				
2. MECHANICAL DESIGN	45				
a. Structural	8				
b. Coatings	8				
c. Rollers	6				
d. Belt Tracking System	6				
e. Bearings	6				
f. Control System	3				
g. Overall Assessment of Machine Integrity	8				
3. VENDORS' EXPERIENCE	20				
a. Total Installations	4				
b. Pulp and Paper Installations	4				
c. U.S. Installations	2				
d. Experience and Knowledge of Belt Press Theory	5				
e. Knowledge of Construction and Design	5				
4. VENDORS' REPUTATION/SERVICE	10				
a. Location of Sales and Service Reps	2				
b. Track Record of Maintenance Requirements	4				
c. Proven Record of Response to Client Needs	4				
5. CONFIDENCE IN VENDOR	20				
a. Response to & Understanding of Project Needs	2				
b. Completeness of Proposal/Submittal Information	2				
c. Performance Warranty	3				
d. Intangibles (personal confidence, reputation, etc.)	3				
e. Vendor Presentation	10				
6. PRICE	50				
a. Bid Price of Machine — Point Distribution					
b. Building Costs — Based on Present					
c. O & M Costs — Worth Analysis					
TOTAL	185				

D. Cost Benefits

As indicated in Table #3, the WLSSD belt press system has exceeded vacuum filtration performance. Improved solids recovery, lower sludge cake volume and higher sludge cake volatile content have contributed to more effective plant operation.

Sludge cake volume reduction has occurred in spite of lower cake dryness produced by the presses. The decrease of 1.26 metric tons cake per ton D.S. processed (1983 compared to 1981) is due to elimination of inorganic conditioning chemicals. Based on WLSSD's annual budget sludge production level of 10,900 tons D.S. the belt presses produced a savings of $80,000 to $100,000 per year in land application and incinerator fuel costs due to sludge volume reduction. Lower ash volumes, also due to elimination of lime and ferric chloride conditioning, produced savings in ash transportation and landfilling.

Yet another benefit was produced by the replacement of four flat belt conveyors with the single Serpentix conveyor. The flat belt units provided with the vacuum filters required constant spraywashing for cleaning, which recycled approximately 45 metric tons D.S. into the plant process per month. The Serpentix unit eliminated this solids recycle completely, reducing dewatering loading, and if calculated on a per ton processed basis, provided an annual savings of about $45,000.

Combined cost benefits of the belt press project have produced the predicted payback period of about one year.

TABLE #3

WESTERN LAKE SUPERIOR SANITARY DISTRICT
PERFORMANCE COMPARISON
VACUUM FILTER TO FILTER BELT PRESS DEWATERING

	VACUUM FILTERS *	BELT PRESSES**
T.P.D. Per Unit	16.5 metric tons D.S.	24.5 metric tons D.S.
% TS Cake	20.3%	16.0%
%TVS Cake	54.9%	70.1%
Fuel Value, Cake	3522 Gm. cal./gm.	3923 Gm. cal./gm.
Metric Tons Cake/Ton D.S.	7.0 Tons	5.74
Metric Tons Ash/Ton D.S.	.67 Tons	.27 Tons
Solids Recovery	93.6%	94.8%

* 1981 VF Operation

** 1983 FBP Operation

Table #4 compares costs of vacuum filtration operation (base period 1981) to belt press operation (base period first nine months 1983), without adjustment for inflation. Costs other than filtration chemicals are those of the entire dewatering department (including the flotation thickening process). Maintenance cost savings were inconclusive as they reflect several modifications and one major one-time repair to press roll coverings.

WESTERN LAKE SUPERIOR SANITARY DISTRICT
PERFORMANCE COMPARISON
DEWATERING DEPT. COSTS, 1981, 1983
VACUUM FILTER TO FILTER BELT PRESS DEWATERING

TABLE #4

	VACUUM FILTERS*	BELT PRESSES**
Filtration Chemicals per metric ton D.S.	$ 53.34	$ 33.33
Power Consumption		
KWH/metric ton D.S.	189 KWH	128 KWH
$ /metric ton D.S.	$ 8.33	$ 5.93
Operations Labor Cost		
Operations Personnel	14	7½
$ /metric ton D.S.	$ 29.58	$ 20.84
Dept. Maintenance Cost $ /metric ton D.S.	$ 5.63	$ 10.13

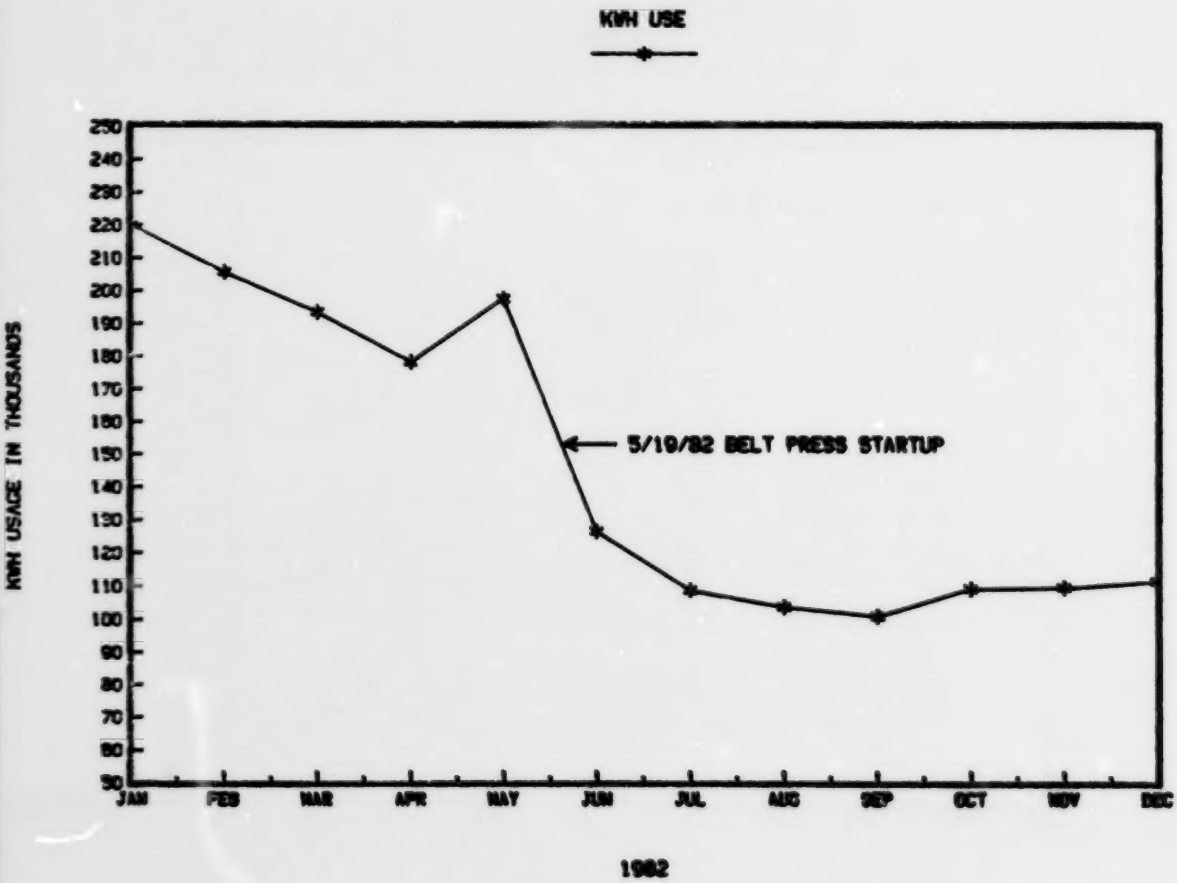
* 1981 VF Operation, 1981 Dollars, 11460.7 metric tons D.S.

** 1983 FBP Operation, 1983 Dollars, 9656.2 metric tons D.S.

The most striking effect of the belt press system was reduction in power consumption in the dewatering department. Replacement of four vacuum filters requiring 400 connected horsepower with two belt presses at 12 connected horsepower, immediately eliminated nearly 100,000 KWH per month in power usage. Monthly total power usages for the dewatering department (including flotation thickening) displayed in Figure #3 indicate this reduction very clearly.

WLSSD DEWATERING ENERGY CONSUMPTION

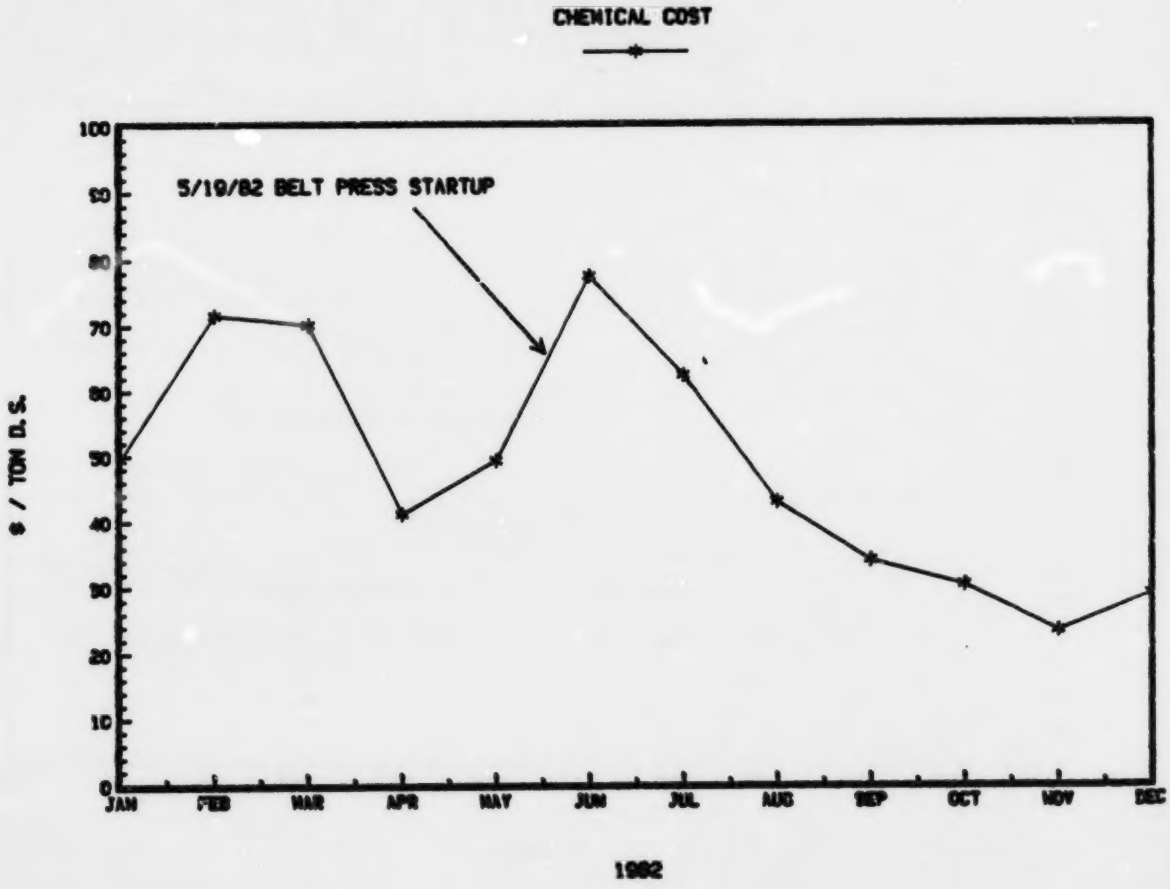
FIGURE #3



Filtration chemical costs were also reduced dramatically by the process change. Figure #4 actually indicates a rise in per ton chemical cost shortly after startup. This was caused by a change in sludge polymer requirements between pilot testing and startup. The polymer identified as the optimum product during the pilot test proved marginally effective shortly after startup, thus producing high per ton processed costs. Change from a dry cationic to a liquid mannic cationic polymer in August 1982 brought chemical costs to lower than expected levels.

WLSSD DEWATERING FILTRATION CHEMICAL COST

FIGURE #4



The most concrete evidence of the success of any cost-saving measure (and most politically visible) is the budget process. Annual budgets for the Dewatering section graphed in Figure #5 show a comfortable trend, largely due to the change in dewatering process.

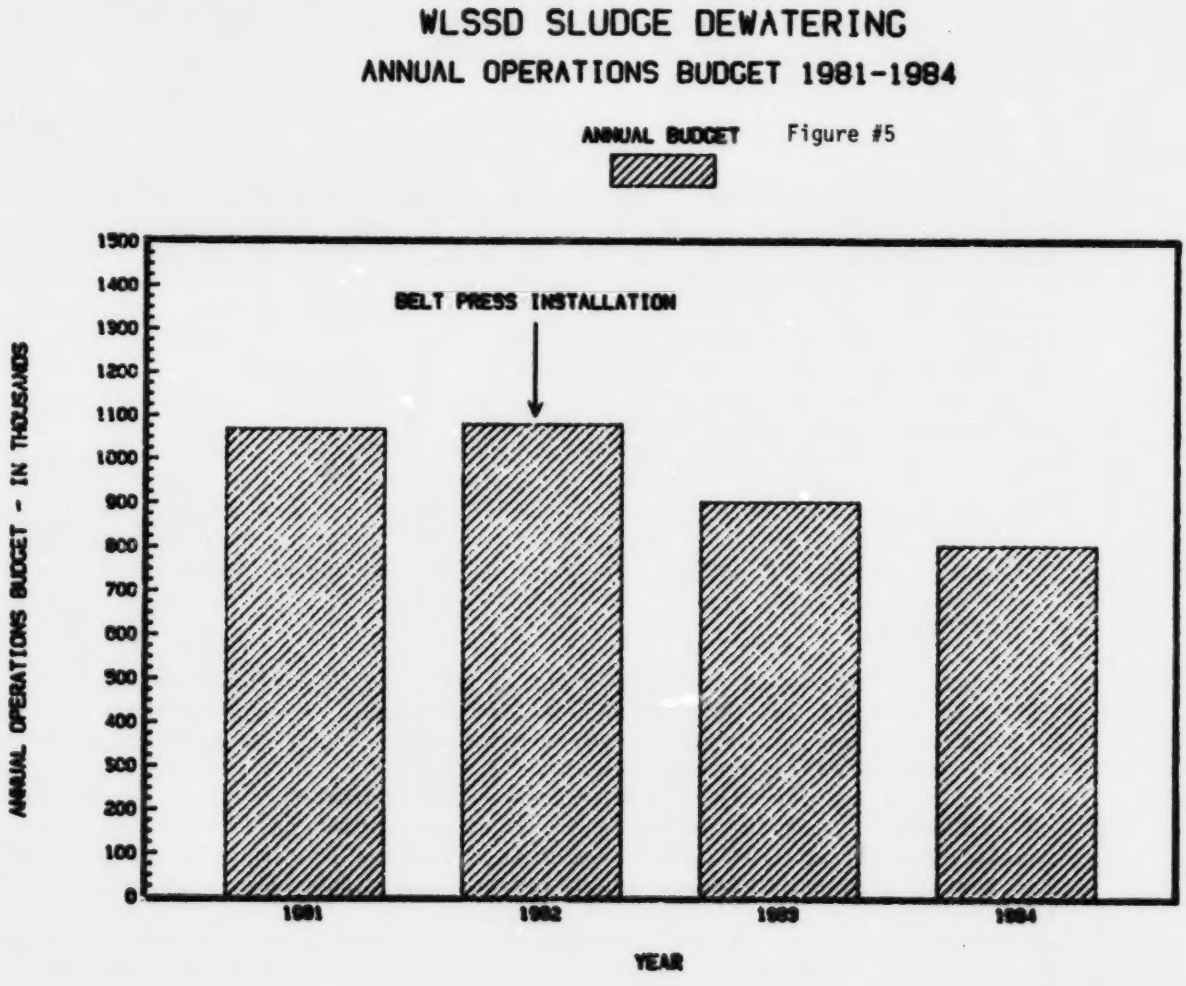


Figure #5

WLSSD's change to filter belt press dewatering produced significant personnel changes. Department staffing was reduced from fourteen to seven and one-half positions, shift coverage of the area was reduced from two to one operators, and two full-time supervisory positions were combined into one position for the entire wastewater treatment plant.

Additional benefits occurred in safety conditions in dewatering. Elimination of lime slaking, ferric chloride usage, acid handling, sludge spillage and drain backups removed hazardous conditions that contributed to sixteen personnel injuries in 1981. Accidents experienced in 1982 and 1983 dropped to six per year, largely due to the belt press installation.

Finally, because the belt presses are compact and require a minimum of support equipment, a small area of vacant space was used for the new installation. This freed up 550 square meters of building space formerly used for vacuum filtration.

Belt Press Performance

The belt presses have been sound mechanically and structurally over twenty months of operation. One major repair was required to replace FRP roll coatings due to abrasion after one year of operation. Replacement of FRP with a 70 durometer rubber coating has proven successful. Three major and about a half-dozen minor roll bearings have been replaced. The press frames and rolls have performed very well with no structural failures.

Press wire life exceeds 2000 operating hours, with 3000 hrs. expected after additional wire optimization. Electronic drives, and pneumatic tracking and tensioning systems have worked well.

The belt presses have consistently met performance minimums in polymer consumption, solids recovery and throughput. In fact, the units have demonstrated the ability to process 160% of the 27 metric ton per day design rate over a 24 hour continuous run. Sludge cake solids, however, have averaged one to two percent below the 18% TS design level during routine operation.

E. Conclusion

As a process change, replacement of vacuum filtration with belt press dewatering at WLSSD has been an unqualified success. Cost benefits were derived in chemical usage, labor and power consumption. Benefits expected in maintenance have been more difficult to assess.

From an engineering viewpoint, the project was also quite successful. The short timeline contract, with high vacuum filtration daily operating costs as justification, required only three-week extension to complete construction. The performance requirements and fifty percent withholding were a useful lever to assure system construction and performance as specified. Turnkey contracting with the supplier as general contractor worked fairly well initially, with rapid flow of information on change orders, design and drawing approvals, and field modifications. Supplier response was poor late in the project, especially in punch list completion and negotiation over non-performance of prethickener units. The contract was finally completed in early 1984 after it was determined that the prethickener units would not meet performance expectations.

Overall, however, WLSSD's experience with belt presses has been more successful than that of many other owners, largely due to the care taken in preliminary investigation, pilot testing, system design and selection of equipment.

AUTHORS

At the time of the filter belt press installation project, Terry Zaudtke was a Staff Engineer for the W.L.S.S.D. and Will Haapala was an Assistant Process Superintendent for W.L.S.S.D. Mr. Zaudtke is now a Project Manager with Conklin, Porter and Holmes Consulting Engineers, Inc. Sanford, Florida, and Mr. Haapala is Wastewater Treatment Plant Superintendent for the City of Little Falls, Minnesota.

UTILISATION AGRICOLE DES FUMIERS ET DES BOUES

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UTILISATION AGRICOLE DES BOUES D'ÉPURATION

Dans le passé, l'épandage des boues d'épuration était de pratique courante et ne faisait pas l'objet de contrôles quant au taux ou quant à la fréquence d'application. La croissance de l'urbanisation et de l'industrialisation ainsi que la sophistication des procédés de traitement des eaux usées ont conduit à une augmentation de la production des boues d'épuration et en ont modifié la composition chimique.

Dès lors, l'utilisation des boues d'usine d'épuration pour fin d'amendement des sols a fait l'objet de plusieurs études visant à déterminer les risques et les bénéficies d'une telle pratique, ainsi que les normes sécuritaires d'application.

Dick et alii ont comparé les coûts de différents modes de gestion intégrée des boues d'épuration; ils ont conclu que l'épandage des boues liquides était le mode d'élimination le plus économique pour des usines d'une capacité allant jusqu'à 38 000 m^3/d (10 Mgal US/d), et ce dans un rayon d'environ 130 km (80 mi). Le transport des boues liquides par camion est économique quant à lui dans un rayon de 30 km.

En Ontario, en 1978, les boues de 63,2% des usines d'épuration étaient épandues sur des terres agricoles, soit 34% en poids de boues produites. La région de Toronto et les villes de Hamilton et de London éliminaient 41% des boues produites en Ontario par incinération, 16% des boues étaient enfouies et 9% éliminées par d'autres moyens. Bien que l'utilisation des boues pour fins d'agriculture gagne la faveur des petites municipalités, le pourcentage des boues épandues est actuellement à la baisse à cause de la construction de nouveaux incinérateurs dans la région de Toronto et de l'application plus stricte des directives ontariennes concernant l'épandage.

L'utilisation de boues liquides pour fins d'agriculture permet d'en préserver la valeur fertilisante; à cet égard, il s'agit donc d'une solution préférable au compostage qui produit des amendements de sol à faible teneur en éléments nutritifs.

Les boues doivent être emmagasinées lorsque les conditions de sol ne permettent pas l'épandage. Au Québec, la durée de l'entreposage peut atteindre six mois.

La réglementation concernant l'épandage des boues d'épuration pour fins d'agriculture est nécessaire pour préserver la qualité des sols et des eaux souterraines ainsi que la qualité des récoltes.

Le Québec est en voie de se doter de directives à ce sujet. Les directives ontariennes ont été publiées pour la première fois en 1973 et ont fait l'objet de plusieurs révisions depuis. Les principaux points visés par ces directives sont les suivants:

a. Seules les boues digérées ou stabilisées sont acceptables pour fins d'agriculture.

b. Les cultures les plus appropriées pour bénéficier de l'épandage des boues d'épuration sont le maïs, les herbages pour foins et pâturages et la tourbe à gazon commerciale.

c. Les terres ne peuvent être utilisées pour la culture des légumes ni comme pâturage pour les moutons et les porcs dans les six mois suivant l'épandage des boues d'épuration. Les chevaux, les bovins et les vaches laitières peuvent accéder à de tels pâturages deux mois après l'épandage. Pour la culture des fruits, il est recommandé d'épandre les boues d'épuration à la fin de l'automne (en novembre) ou au plus tard trois mois avant la récolte.
d. Les boues d'épuration ne doivent pas être épandues sur des terres organiques ni sur des sols contenant plus de 60 ppm de phosphore disponible.
e. Le taux d'application d'azote ne doit pas excéder les besoins des cultures ou au maximum 135 kg/ha d'azote ammoniacal et de nitrates (NH_4^+ -N et NO_3^- -N) pour une période de cinq ans. Dans le cas de la tourbe à gazon, la période de récurrence tolérée est de quatre ans.
f. Des limites absolues sont établies pour les taux d'application de onze métaux, ainsi que des limites relatives au contenu des boues en azote.
g. Les boues d'épuration ne peuvent pas être épandues sur des sols ayant un pH inférieur à 6,0, à l'exception des boues de chaux qui peuvent être épandues sur des terres acides.
h. Les boues ne doivent pas être épandues sur des terres où le roc est à une profondeur inférieur à 1,5 m à moins qu'il ne soit prouvé que les risques de contamination des eaux souterraines et de surface soient négligeables.
i. Il est recommandé qu'un épandage unique ne dépasse pas 130 m^3/ha (1,3 cm d'épaisseur).

Faisabilité de l'utilisation agricole des boues d'épuration dans la région de Montréal

Une étude préliminaire a été menée pour évaluer la faisabilité de l'utilisation agricole des boues dans la région de Montréal. Les municipalités faisant partie de la phase I du programme d'assainissement du gouvernement du Québec ont fait l'objet de cette étude, à l'exception de la communauté urbaine de Montréal. La superficie requise pour l'épandage des boues d'épuration produites par ces populations a été calculée. D'autre part, le dénombrement des populations animales de ces régions a servi à évaluer la superficie requise pour l'épandage des fumiers. La superficie susceptible d'être utilisée pour l'épandage a été délimitée au moyen de méthodes cartographiques afin d'établir la distribution possible des besoins de chaque municipalité.

Le tableau 1 résume les données et les résultats obtenus. La distance maximale de transport des boues varie de 8 à 30 km pour les municipalités considérées. Il est évident que la faisabilité de l'utilisation agricole des boues est d'autant plus grande que cette distance est plus petite. En première approche, il semblerait donc possible d'envisager ce mode de gestion des boues dans un bon nombre de ces cas.

Faisabilité de l'utilisation agricole des boues d'épuration de la ville de Châteauguay

La ville de Châteauguay étant située dans une région agricole privilégiée, les autorités municipales prévoient recourir à l'utilisation des boues d'épuration en agriculture.

TABLÉAU 1 - SUPERFICIES REQUISES POUR L'EPANDAGE DES BOUES D'EPURATION ET DES FUMIERS - REGION DE MONTREAL

1	2	3	4	5	6	7	8	9	10	11	12	13
MUNICIPALITES	POPULATION	INDUSTRIES: POPULATION EQUIVALENTE	POPULATION TOTALE (2+3 x1.2)	SUPERFICIE REQUISE (13 p./ha)	NOMBRE DE PORCS (+20%)	EQUIVALENT UNITES ANIMALES (5/U.A.)	NOMBRE DE BOVINS (+20%)	EQUIVALENT UNITES ANIMALES (variable)*	UNITES ANIMALES TOTALES (7 + 9)	SUPERFICIE REQUISE (.3 ha/UA)	SUPERFICIE TOTALE (ha) (5 + 11)	DISTANCE MAX. DE TRANSPORT DES BOUES
Boisbriand	11 000	5 294	19 553	1 504			14 880	V_1 14 880				
St-Eustache	22 300	--	26 760	2 058	10 680	2 136	2 760	V_2 1 380	18 636	5 591	10 002	11 km
Deux-Montagnes	9 200	--	11 040	849			2 400	V_{10} 240				
Beauharnois							12 840	V_1 12 840				
Maple Grove	7 800	--	9 360	720	1 320	264	3 000	V_2 1 500	14 808	4 442	5 162	11 km
							2 040	V_{10} 204				
Châteauguay	50 000	--	60 000	4 615	5 520	1 104	19 080	V_1 19 080				
							4 680	V_2 2 340	22 836	6 851	11 466	16 km
							3 120	V_{10} 312				
Ste-Catherine												
St-Constant							6 000	V_1 6 000				
Delson	34 200	--	41 040	3 157	14 520	2 904	1 080	V_2 540	9 504	2 851	6 008	8 km
Candiac							600	V_{10} 60				
Laprairie												
St-Hyacinthe	43 900	88 236	158 563	12 197	77 520	15 504	18 960	V_1 18 960				
							4 200	V_2 2 100	36 924	11 077	23 274	19 km
							3 600	V_{10} 360				
Farnham	6 500	--	7 800	600	32 160	6 432	19 560	V_1 19 560				
							4 560	V_2 2 280	28 656	8 597	9 197	14 km
							3 840	V_{10} 384				
Cowansville	12 000	3 824	18 989	1 461	39 000	7 800	27 000	V_1 27 000				
							6 240	V_2 3 120	38 656	11 567	13 028	30 km
							6 360	V_{10} 636				
Acton Vale	4 234	16 659	25 072	1 929	269 160	53 832	29 520	V_1 29 520				
							6 720	V_2 3 360	87 384	26 215	28 144	27 km
							6 720	V_{10} 672				
Granby	37 000	57 918	113 902	8 762	124 080	24 816	33 840	V_1 33 840				
Waterloo	4 700	1 070	6 924	533			6 120	V_2 3 060	62 424	18 727	28 022	30 km
							7 080	V_{10} 708				
Laval	239 200	--	287 040	22 080	1 200	240	1 440	V_1 1 440				
							120	V_2 60	1 764	529	22 609	24 km
							240	V_{10} 24				

* V_1 signifie un rapport de 1/1
V_2 signifie un rapport de 1/2
V_{10} signifie un rapport de 1/10

Une étude a été menée dans les environs immédiats de la ville de Châteauguay pour vérifier la disponibilité des sols agricoles propices à l'épandage des boues d'épuration, sensibiliser le milieu professionnel agricole à cette technique et recueillir ses réactions.

Les zones non propices à l'épandage ont été identifiées, à savoir:

a. Les superficies boisées, non pas que les boues soient dommageables aux boisés mais plutôt à cause des difficultés techniques de l'épandage. Il faudrait en effet y utiliser un système d'irrigation ou y multiplier les chemins d'accès, solutions qui, à ce stade, ont été rejetées.

b. Les zones urbaines et para-urbaines comprenant les secteurs résidentiel, commercial et industriel.

c. Les zones d'extraction de minerais, de pierre, de sable, de gravier, de tourbe et de sol arable.

d. Les secteurs de loisirs et les chalets.

e. Les terres improductives, soit les terres marécageuses, le sable nu, le roc à nu et toute autre terre utilisée à d'autres fins que l'agriculture.

f. Les terres non exploitées; friches à couverture herbacée et broussailles.

Parmi les terres propices, certaines ont été écartées à cause des cultures qui y sont pratiquées, soit l'horticulture ou des cultures spéciales telles que pommes de terre, betteraves sucrières, tabac, pépinière, bleuets indigènes, graines oléagineuses ou vergers en production. L'épandage sur ces terres comporte des limitations et exigerait une gestion élaborée.

En procédant par élimination, il a donc été possible de délimiter les terres agricoles propices, soit celles où se pratiquent la grande culture et les pâturages améliorés.

En ne considérant que les terres situées dans un rayon de 16 km (10 mi) au sud de la future usine d'épuration, la superficie des terres agricoles propices à l'épandage des boues d'épuration a été évaluée à 13 300 ha (32 850 acres).

Un inventaire réalisé par le Bureau régional du ministère de l'Agriculture du Québec en 1981 révèle la présence de 8418 unités animales pour les municipalités de Châteauguay, Mercier, St-Urbain, Ste-Martine, St-Rémi et St-Isidore. Sur la base de 0,3 ha par unité animale pour l'épandage du fumier, 2525 ha (6250 acres) de terres doivent être réservées pour les exploitations animales.

Étant donné que la distribution de la population animale dans le secteur n'était pas connue, toute la superficie calculée a été déduite. La superficie nette disponible et propice sans limitation pour l'épandage des boues d'épuration est donc évaluée à 10 775 ha (26 600 acres).

L'évaluation des superficies requises pour l'épandage se fait ensuite à partir d'un taux maximal d'application d'azote assimilable de 130 kg/ha - 5 ans.

L'azote assimilable dans les boues d'épuration est évalué à 2,5% des matières solides totales pour le type de traitement prévu à Châteauguay.

Pour un volume annuel de boues de 20 000 m^3/an contenant 730 000 kg de matières solides, la superficie requise est donc calculée comme suit:

$$730\ 000\ kg \times \frac{2,5\%N}{130\ kg/ha} = 140\ ha$$

Comme cette dose assimilable ne peut être répétée qu'à tous les cinq ans, il faut prévoir une banque de terre cinq fois plus grande, soit 700 ha. Cette banque ne représente que 6,5% des terres disponibles dans le rayon de 16 km préétabli.

Dans le cadre de cette étude, plusieurs personnes du milieu agricole ont été interviewées. Il s'agit d'agronomes et d'ingénieurs du ministère de l'Agriculture du Québec, de conseillers en génie rural ou en grandes cultures, de représentants de l'union des producteurs agricoles, d'opérateurs d'usines d'épuration, d'agriculteurs et de représentants des grandes conserveries de la région.

De l'ensemble de ces entrevues, les observations suivantes peuvent être extraites:

a. Les boues d'épuration ne sont pas connues ou très peu. On sait que l'épandage sur sols agricoles se pratique, mais sans savoir selon quelles modalités.

b. Les agriculteurs sont intéressés à recevoir les boues à condition que leur valeur fertilisante soit connue et que leur épandage ne cause pas d'ennuis.

c. Les professionnels manifestent un grand intérêt malgré une pointe d'incertitude, notamment face aux métaux et à leur contamination possible des sols et des plantes.

d. Les producteurs de maïs sont les plus intéressés, à cause de la valeur fertilisante et de l'amendement fourni par les boues.

e. L'absence de règles ou d'un guide québécois sur le sujet semble manquer à plusieurs, notamment aux professionnels.

En conclusion, la région de Châteauguay est un milieu favorable à l'utilisation agricole des boues d'épuration puisque les cultures céréalières et commerciales y abondent et qu'il n'existe pratiquement pas de compétition de la part des fumiers provenant d'exploitations animales.

Cependant, aucune municipalité québécoise de l'envergure de Châteauguay ne pratique encore l'épandage des boues d'épuration. L'implantation de cette technique constituera donc une première et devra être abordée comme telle par la mise en place de procédures permettant d'assurer une bonne gestion agricole des boues d'épuration.

PROBLÉMATIQUE GÉNÉRÉE PAR L'ABONDANCE DES LISIERS DE PORCS AU QUÉBEC

Au Québec, l'une des principales objections que l'on rencontre face à l'utilisation agricole des boues d'épuration provient de la surabondance des lisiers de porcs dans certaines régions et des importants problèmes de pollution qui y sont associés.

Au cours de l'année 1983-84, un projet pilote de gestion de l'épandage de lisiers, subventionné par Environnement Canada, a été réalisé par PELLEMON, en étroite collaboration avec le service d'assainissement agricole du ministère de l'Environnement du Québec. Ce projet atteint les principaux objectifs suivants:

- Concevoir un outil cartographique régional pour la gestion des lisiers.
- Expérimenter un programme de gestion des lisiers entre éleveurs et agriculteurs dans le cadre d'un projet pilote desservant une population porcine limitée.
- Établir le coût, les avantages et les inconvénients du transport et de l'épandage par injection des lisiers.
- Fertiliser avec des lisiers à l'intérieur d'un programme de fertilisation déterminé.

En cours de réalisation, et bien que cela ne constituait pas un objectif du projet original, il a été nécessaire de faire une analyse de la conception mécanique des équipements employés.

Cartographie d'une région problème

L'outil cartographique développé est en quelque sorte un plan directeur de gestion des lisiers dans le bassin versant de la rivière L'Assomption. Ce bassin regroupe les plus fortes populations porcines au nord de Montréal, St-Roch-de-L'Achigan se classant au premier rang avec plus de 150 000 porcs.

La délimitation du territoire s'est effectuée en entourant chacune des municipalités à forte population porcine d'une ceinture de municipalités tampons capables d'assimiler les surplus de lisiers. Ainsi, le territoire étudié comporte les municipalités suivantes:

Dans le comté de Joliette: St-Jean-de-Matha, Ste-Béatrix, St-Alphonse-de-Rodriquez, St-Félix-de-Valois, Ste-Mélanie, Ste-Marcelline-de-Kildare, St-Ambroise-de-Kildare, Notre-Dame-de-Lourdes, Notre-Dame-des-Prairies et Joliette.

Dans le comté de Montcalm: Rawdon, St-Calixte, St-Liguori, St-Jacques, St-Alexis, Ste-Julienne, Ste-Marie-Salomé et St-Esprit.

Dans le comté de L'Assomption: St-Roch-de-L'Achigan, St-Lin, St-Gérard-Majella, L'Épiphanie, L'Assomption, Laplaine, Mascouche et Le Gardeur.

Dans le comté de Terrebonne: Ste-Anne-des-Plaines.

Suivant les derniers recensements, les populations animales de ce territoire s'élèvent à 115 387 unités animales dont 69 623 unités animales de porcs et de truies soit près de 350 000 bêtes.

La superficie requise pour utiliser tous les fumiers produits dans ce territoire est évaluée à 34 619 ha tandis que la superficie cultivable s'élève à 63 250 ha.

Il est donc nécessaire d'épandre des fumiers sur 55% des superficies en culture pour disposer de la totalité des fumiers du territoire. Dans une telle hypothèse, l'analyse de l'autosuffisance de chacune des municipalités permet d'identifier celles qui ont un déficit de superficies cultivables par rapport à leur production de fumiers et qui devront acheminer leurs surplus à l'extérieur de leurs frontières. Il s'agit des municipalités de St-Roch-de-L'Achigan et de St-Liguori (déficit 7769 ha et 1197 ha), St-Jean-de-Matha (267 ha), St-Ambroise-de-Kildare (210 ha), St-Alexis (28 ha), St-Esprit (1384 ha) et St-Lin (216 ha).

La méthodologie de gestion proposée consiste à acheminer les surplus de fumiers des municipalités déficitaires vers les municipalités périphériques où sont disponibles des terres d'épandage, tout en veillant à minimiser les distances à parcourir.

L'attribution des superficies d'épandage se fait en débutant par les municipalités les plus distantes du point le plus critique. Cette séquence évite de déplacer le problème, c'est-à-dire d'assigner des terres éloignées à une municipalité n'ayant que peu de surplus de fumiers, alors que des superficies d'épandage en bordure de ses limites sont disponibles mais attribuées à une municipalité plus problématique.

Cet exercice montre qu'il est nécessaire de parcourir 36 km à vol d'oiseau pour décongestionner le cas le plus critique, soit St-Roch-de-L'Achigan dont une partie des surplus seront acheminés à Ste-Marcelline-de-Kildare et Notre-Dame-de-Lourdes.

Cependant, ce parcours, mesuré à vol d'oiseau, fait abstraction des contraintes physiques imputables au réseau routier de la région. Par conséquent, une évaluation plus précise du kilométrage nécessite la localisation des terres en culture, de même que la détermination des routes empruntées pour le transport des lisiers. Pour ce travail, des cartes cadastrales à l'échelle 1:20000 ont servi de plans de base. En plus des réseaux hydrographique et

routier, ces cartes permettent la localisation et la délimitation exacte des terres. Il résulte de ce travail détaillé que la distance de 36 km à vol d'oiseau correspond à un parcours de 48 km par les routes.

L'étude de la répartition des fumiers par rayon d'épandage permet d'établir que si l'épandage se fait sur 55% des terres cultivées, 50% de tous les fumiers seront épandus à l'intérieur d'un rayon de 7,8 km, 95% à l'intérieur d'un rayon de 28 km, tandis que l'épandage des 5% excédentaires nécessitera un parcours susceptible d'atteindre 48 km.

Expérimentation d'un programme de gestion

Le projet était consacré à l'épandage par injection en postlevée. Cette méthode consiste à soulever la terre entre les rangs de culture, quand les plants sont visibles, et à injecter le lisier dans le sol et non en surface. La durée de l'épandage par cette méthode est limitée par la croissance des plants et peut se poursuivre pendant environ six semaines.

Les premières activités du projet pilote ont été dévolues à la préparation des opérations d'épandage:

- Sélection des fermes d'élevage et d'épandage et visite de chacun des agriculteurs et des éleveurs invités à participer au projet.
- Assemblée d'information des participants.
- Planification du programme d'épandage.
- Choix des routes, calendrier, calculs de fertilisation.
- Échantillonnage des sols et des eaux de lixiviation.
- Conception des formulaires d'opération.
- Préparation des équipements d'épandage:
  . Un tracteur routier dix (10) roues;
  . Deux citernes semi-remorques d'une capacité nominale de 36 m^3 chacune (capacité utile 27 m^3);
  . Un tracteur de ferme John-Deere quatre (4) roues motrices modèle 3140;
  . Une citerne de ferme de 10 m^3 de capacité nominale (capacité utile 9 m^3) avec quatre (4) roues sur tandem à voie ajustable de 1524 mm (60 po) à 3658 mm (144 po);
  . Un Purinject Lajoie à deux injecteurs ajustables de 1524 mm (60 po) à 1829 mm (72 po) d'espacement;
  . Une pompe de transfert 76 mm ϕ (3 po), 30 CV diesel sur roues avec boyaux à raccords rapides.

Analyse des considérations agricoles. La valeur fertilisante du lisier utilisé pour le projet a été établie au moyen d'analyses à l'hydromètre de Tunney. Cet instrument permet d'estimer la valeur fertilisante du lisier en fonction des taux de matières sèches. Des analyses en laboratoire ont également été faites en vue de jauger l'hydromètre avec des lisiers produits au Québec.

Le calcul des doses d'épandage se base sur les hypothèses suivantes:

- Le facteur d'utilisation est établi en fonction du type de sol:
  . Argile: 0,80
  . Loam: 0,75
  . Sable: 0,65
- Seul l'azote assimilable est considéré, soit 70% de la valeur obtenue à l'hydromètre.
- Le dosage est calculé selon les besoins en azote seulement, les autres éléments fertilisants sont négligés.
- Le dosage tient compte des autres engrais utilisés par l'agriculteur.

Avec l'agriculteur, il était décidé de remplacer une proportion définie des engrais azotés qu'il prévoyait utiliser. D'une façon générale, le remplacement ne dépassait pas 50% des besoins.

À partir du dosage d'azote à l'hectare peut être calculée la quantité de lisier à épandre à l'hectare selon sa qualité. Il reste ensuite à établir le trajet à parcourir avec l'épandeur en fonction de l'espacement entre les rangs et de la longueur de la parcelle.

L'expérience démontre cependant que, dans le cas d'injection en postlevée, les contraintes physiques de la parcelle et du volume de lisier contenu dans la citerne fixent pratiquement le dosage. En effet, puisqu'il est impossible de quitter le rang en plein milieu d'un champ en postlevée dans des cultures en rangées, la citerne est vidangée en un nombre de passages fixes et la dose réelle de lisier épandu varie en proportion des fluctuations dans la qualité du lisier.

L'injection dans les sols sableux (ou autres sols légers) a été réalisée sans difficulté. Ce type de sol camoufle aisément le lisier et l'absorbe rapidement tout en produisant une texture granuleuse salutaire aux cultures.

Par contre, l'injection dans les sols lourds (argileux) s'est avérée problématique à cause de la géométrie du socle d'injection, mais également à cause des propriétés intrinsèques des sols eux-mêmes qui n'absorbent que très lentement le lisier et le dissimulent mal.

Le stade de croissance des plantes au moment de l'injection joue également un rôle dans les résultats obtenus. Lorsque leur croissance est peu avancée, la terre déplacée par le passage de l'injecteur déplace ou déracine certains plants, mais ceux-ci reprennent facilement. Par contre, lorsque la croissance est plus avancée, moins de plants sont endommagés puisqu'ils sont mieux enracinés; toutefois, ceux qui sont déracinés subissent trop de dommages pour atteindre la maturité.

Dans les champs à couverture herbacée, le travail de l'injecteur est remarquable, le lisier étant injecté sous la couche végétative qui se referme par la suite sans laisser échapper de liquide.

Dans toutes les situations où l'injection a été complète et satisfaisante, aucune odeur notable de lisier ne pouvait être sentie. Il s'agit là d'un élément positif majeur; de l'injection pourrait être avantageusement effectuée à proximité des zones habitées, même en été par période de temps chaud.

Équipements. Les principales difficultés mécaniques rencontrées sont les suivantes:

a. Tracteur. La citerne de lisier étant lourde, le tracteur doit être suffisamment lesté à l'avant pour assurer un bon équilibre des masses. L'emploi d'un tracteur à quatre (4) roues motrices s'est révélé avantageux.

b. Citerne. Plusieurs éléments de la citerne se sont montrés défaillants ou trop faibles pour l'usage auquel ils étaient destinés:

- Perforations dans le corps de la pompe arrière;
- Ouvertures d'échappement d'air absentes;
- Robinets et vannes de contrôle de débit trop fragiles;
- Roulement à bille de l'arbre d'entraînement entre le tracteur et la citerne de calibre trop faible et mal aligné.

D'autre part, il s'est avéré difficile de s'aligner entre les rangs sans piétiner des plants et les manoeuvres de virage aux extrémités des champs étaient compliquées par les ponceaux et les barrières qui s'y trouvaient.

c. Injecteurs. Les coutres étaient trop fragiles et auraient gagné à être gaufrés, beaucoup plus grands (600 mm de diamètre au lieu de 380), à être fixés plus solidement au bâti et à être munis d'une protection de relèvement à ressort ou par boulon à déchirement plutôt que le mécanisme de tiges et cylindres concentriques utilisé. Les bras de soutien des injecteurs étaient également trop faibles.

Dans les sols lourds, l'injection a déplacé de grosses mottes de terre, laissant une tranchée ouverte à l'arrière du socle. Par conséquent, le lisier demeurait à découvert ou les mottes tombaient sur les plants, les endommageant.

Dans les champs où la culture de l'année précédente avait laissé des résidus, des tiges sèches s'accumulaient à l'avant de l'injecteur et sur le tube de support d'alimentation du lisier.

Cet amas de tiges accumulant de la terre, l'injecteur pivotait et se soulevait déposant le lisier sans l'injecter. Le débalancement ainsi produit rendait difficile le contrôle des équipements. Le problème était particulièrement important sur les retours de maïs-grain.

d. Tracteur routier et semi-remorques. Les principales difficultés rencontrées avec ces équipements sont dues aux aménagements des abords des fosses à lisiers et des champs qui ne sont pas aménagés pour recevoir des semi-remorques. Les espaces de manoeuvre sont trop exigus, les entrées de chemin et les ponceaux sont trop étroits, les capacités portantes des sols et des ponceaux sont parfois insuffisantes.

Par ailleurs, les semi-remorques doivent être équipées d'agitateurs permettant de maintenir en suspension les solides du lisier pour éviter qu'ils se déposent et rendent la vidange difficile et lente.

e. Systèmes d'agitation, de reprise et de transfert. La manipulation des boyaux entre les semi-remorques et la citerne d'épandage a provoqué inévitablement de légers déversements de lisier sur le sol. Des améliorations aux boyaux et au système de raccords rapides doivent être apportés pour éviter tout déversement. Idéalement, chaque semi-remorque devrait comporter son système autonome de pompage et de vidange.

Rendement des équipements. Les temps moyens de reprise d'épandage, de chargement et de transfert, de même que les vitesses moyennes de transport des lisiers expérimentés au cours du projet pilote sont présentés au tableau 2.

Il est important de souligner que la vitesse moyenne de déplacement des semi-remorques de lisiers ne peut pas dépasser 35 km/h compte tenu de la circulation et de la topographie des routes empruntées pour des parcours de moins de 50 km.

Coûts de reprise, de transport et d'injection. L'étude du seuil de rentabilité du transport par semi-remorques par opposition au transport par citerne d'épandage démontre qu'il est avantageux de recourir aux semi-remorques à partir d'une distance de 5,7 km entre la fosse à lisier et le champ d'épandage.

Le transport par semi-remorques occasionne cependant du temps d'attente puisqu'il arrive que l'épandage du contenu de la semi-remorque déjà sur les lieux ne soit pas terminé lorsque la semi-remorque pleine arrive au champ. Inversement, il se produit que l'épandage soit déjà complété depuis un certain temps et que l'équipe d'épandage ait à attendre l'arrivée de la semi-remorque pleine. Dans le cas du projet pilote, le temps d'attente était nul pour une distance de 15 km entre la fosse et le champ d'épandage. Il en résulte que

TABLEAU 2 - RENDEMENT COMPARATIF DES ÉQUIPEMENTS

ITEMS	ÉPANDAGE PAR INJECTION TRANSPORT AVEC SEMI-REMORQUES				ÉPANDAGE PAR INJECTION TRANSPORT AVEC CITERNE			ÉPANDAGE PAR AÉROASPERSION TRANSPORT AVEC CITERNE		
	REPRISE	ÉPANDAGE	TRANSPORT		REPRISE	ÉPANDAGE	TRANSPORT	REPRISE	ÉPANDAGE	TRANSPORT
			PELLEMON	MGAF*						
TEMPS MOYEN (min)	11,5	18,5	-	-	6,5	18,5	-	6,5	18,5	-
TEMPS FIXE chargement + transfert (min)	-	38,0	38,0	13,5	-	-	-	-	-	-
VITESSE D'OPÉRATION (km/h)	-	5,0	35,0	80,0	-	5,0	10,0	-	10,0	10,0

* Manuel de gestion agricole des fumiers, MAPAQ.

les coûts de transport sont fixes de 0 à 15 km et équivalents au transport à 15 km, tandis que les coûts d'épandage, qui sont fixes de 0 à 15 km, augmentent au delà de cette distance pour tenir compte du temps d'attente.

La figure 1 illustre l'ensemble de ces coûts.

À partir des coûts déterminés suite à l'expérience réalisée, il est intéressant d'évaluer le coût global d'un programme de gestion à l'échelle de tout le territoire considéré.

Le tableau 3 résume cette analyse économique. Ainsi, l'injection de la totalité des lisiers de porcs de la région pour un scénario d'épandage sur 55% des terres en culture coûterait environ 5,4 M$/année.

Il y aurait lieu d'établir la comparaison avec d'autres modes de traitement des lisiers afin de déterminer un seuil de rentabilité en fonction des distances à parcourir.

CONCLUSION

L'utilisation agricole ne constitue pas l'unique solution au problème d'élimination des boues d'épuration, pas plus que la méthode d'injection ne peut à elle seule éliminer tous les surplus de lisiers. Il s'agit là cependant d'une partie importante du casse-tête que l'on ne peut pas se permettre de négliger. Par ailleurs, ces méthodes sont plus exigeantes que l'usage exclusif d'engrais chimiques, aussi faut-il s'efforcer de faire connaître toutes leurs valeurs additionnelles pour en favoriser l'application.

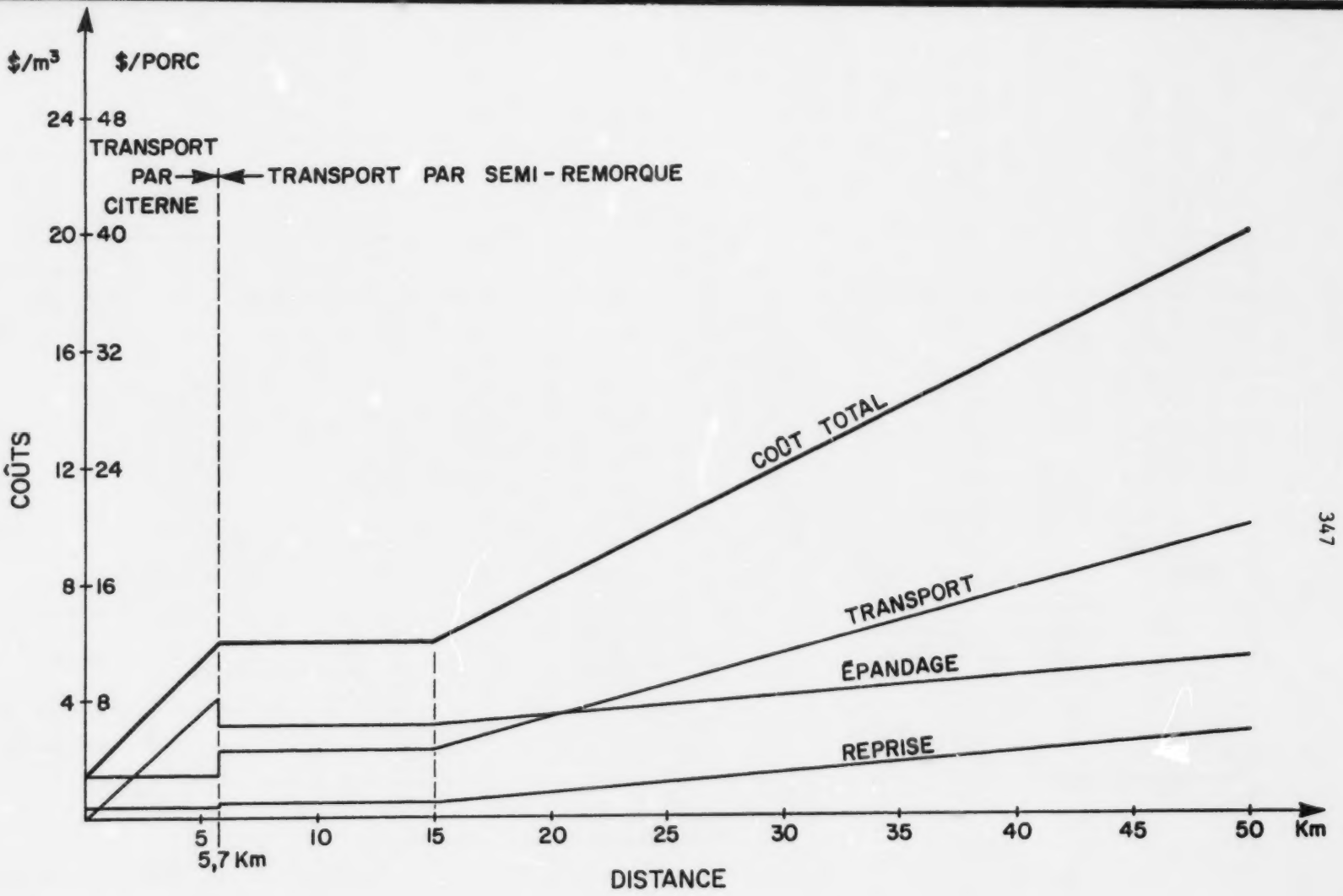

FIGURE 1 - COUT D'EPANDAGE PAR INJECTION (1984)

TABLEAU 3 - RÉGION DE ST-ROCH-DE-L'ACHIGAN
COUT ANNUEL DE L'INJECTION DE TOUS LES LISIERS (1984)

KILOMÉTRAGE		COUT						
	MOYEN	POPULATION PORCINE (u.a.)*	REPRISE ($/m³)	ÉPANDAGE ($/m³)	TRANSPORT ($/m³)	COUT TOTAL ($/m³)	COUT TOTAL ($/u.a.)*	($/an) CUMULATIF
0- 8	5	39 950	0,55	2,35	3,10	6,00	60,00	2 397 000
9-16	12	4 771	0,55	2,35	3,10	6,00	60,00	2 683 300
17-20	19	8 375	0,80	3,36	3,55	7,71	77,10	3 329 000
21-24	23	4 113	1,06	4,45	4,03	9,54	95,40	3 721 400
25-28	27	7 883	1,31	5,55	4,50	11,36	113,60	4 616 900
29-48	40	4 531	2,13	8,99	6,03	17,15	171,50	5 394 000

* unité animale

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CONVERSION OF SEWAGE SLUDGE TO LIQUID FUEL

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INTRODUCTION

Sewage sludge is an unavoidable by-product of wastewater treatment and roughly one tonne of sludge is generated per 4 540 m^3 (1.0 million gallons) of wastewater treated(1). Disposal of sewage sludge is an expensive process and normally constitutes up to 50% of the total annual costs for wastewater treatment. It is estimated that over 500 000 tonnes of dry sewage sludge are produced annually in Canada and disposed of at a total cost of about $104 million(2). Major sludge disposal options used in Canada include agricultural utilization, landfilling and incineration, with disposal costs ranging from about $126/tonne for agricultural utilization to over $300/tonne for incineration(3).

In addition to disposal costs, other considerations such as a projected doubling of sludge production in the next decade(1,4), potential restrictions on agricultural utilization, and an ever increasing difficulty in licensing disposal sites, indicate the need for alternative resource recovery oriented solutions.

TECHNOLOGY REVIEW

Numerous sludge processing options have the potential to convert a fraction of the organic material in sewage sludge into usable energy, but only a few have been demonstrated to be viable net energy producers at full scale. Processes such as starved air incineration, gasification and liquefaction are in development stages and a more thorough review of these is presented elsewhere(5).

The most promising technology has been developed in Germany where researchers from Tubingen University have demonstrated a very simple process to convert sewage sludge to oil and char(6). The process comprises heating dried sludge to 300-350°C in an oxygen free environment for about 30 minutes. Significant advantages of this process are that it operates at only slightly above atmospheric pressure and that no additives are required. The researchers postulated that vapour phase catalysed reactions convert the organics in sludge to straight chain hydrocarbons. The German researchers have demonstrated oil yields ranging from 18-27% and char yields from 50-60%. The oil had a heating value of approximately 39 MJ/kg and the char about 15 MJ/kg. Energy balance calculations indicated the process was a net producer of energy provided that the sludge was mechanically dewatered to about 20% solids. It has been estimated that a net energy production of 7-10 MJ/kg sludge could be demonstrated at full scale.

Environment Canada contracted with Dearborn Environmental Consulting Services to assess this technology in the Canadian context. A two-phase program was developed. The first phase comprised batch experimentation to validate the German process with Canadian sludges and to generate some

limited process design information. The second phase involved construction of a lab-scale continuous reactor system. It was used to generate process design information, for the development of cost data to assess the potential of the technology at full-scale. The final phase of the program will be demonstration of the technology at full-scale.

EXPERIMENTAL METHODS AND TECHNIQUES

Raw sludges containing a mixture of primary and waste activated sludge (WAS) from sewage treatment plants in Toronto, Hamilton and Burlington, Ontario were evaluated during this study. All sludges were oven dried (70°C) to about 90-95% solids. Analyses of these sludges indicated they were similar in composition with volatile solids of 55-64%, calorific values of 15-18 MJ/kg and carbon contents of 24-36% (Table 1).

TABLE 1. SLUDGE ANALYSES (dry weight basis)

	Batch Runs			Continuous Runs
Parameter	Toronto	Hamilton	Burlington	Hamilton
Volatile Solids (%)	60.3	60.5	63.6	55.1
Calorific Value (MJ/kg)	16.0	17.5	17.8	15.05
Carbon (%)	33.28	34.44	36.34	23.8
Hydrogen (%)	4.91	5.17	5.31	3.34
Nitrogen (%)	3.44	3.15	3.63	2.32
Sulphur (%)	0.78	0.96	0.68	1.25
Oxygen (%)	25.41	24.97	28.95	19.2
Nickel (µg/g)	63	222	12	218
Copper (µg/g)	863	655	668	639
Zinc (µg/g)	970	2 416	1 299	2 205
Chromium (µg/g)	1 260	1 260	129	1 540

The batch reactor, a pyrex tube 70 mm in diameter and 720 mm in length, was heated in a three-zone Lindberg furnace under a nitrogen atmosphere (Figure 1). Off-gases were condensed in a trapping system as shown, using ice as the coolant. Non-condensable gases (NCG) were vented from the system. A run was conducted by charging 550 g of dried sludge into the reactor and deaerating with nitrogen. Once operating temperature had been reached, the nitrogen purge rate was reduced. When all visible signs of reaction (i.e., gas/oil flow) ceased, the heat was switched off and the nitrogen purge rate was increased for approximately 30 minutes. The system was dismantled and the char, oil and pyrolytic water collected and stored for analysis. Oil/water separation was achieved using a separatory funnel.

The continuous reactor system is shown schematically in Figure 2. It comprises a stainless steel shell, 5 cm internal diameter by 100 cm long, fitted with a sludge/char conveying system and is heated using the same furnace as for the batch reactor system. The reactor is subdivided by a

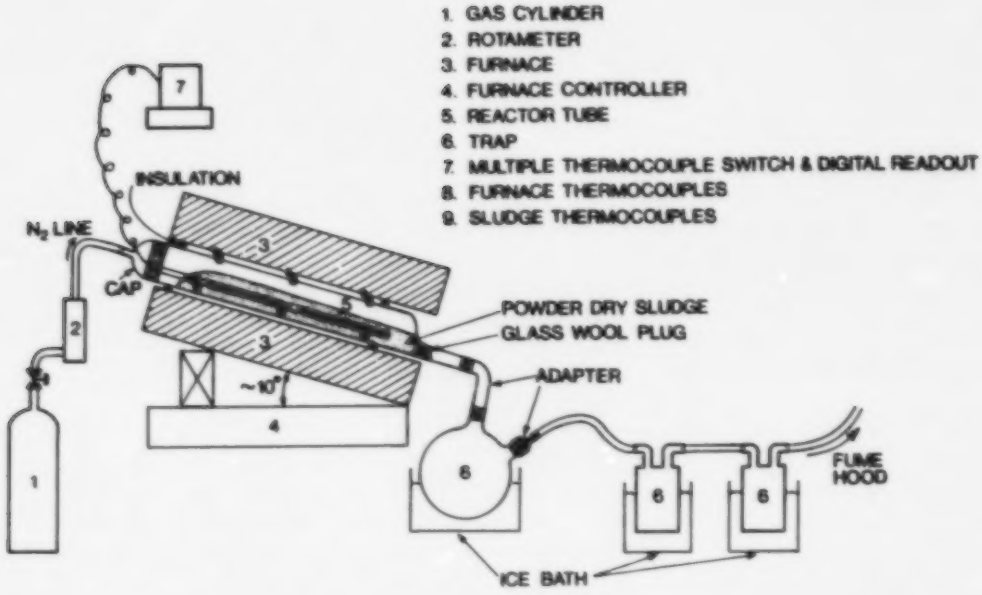

Figure 1. Batch experimental equipment

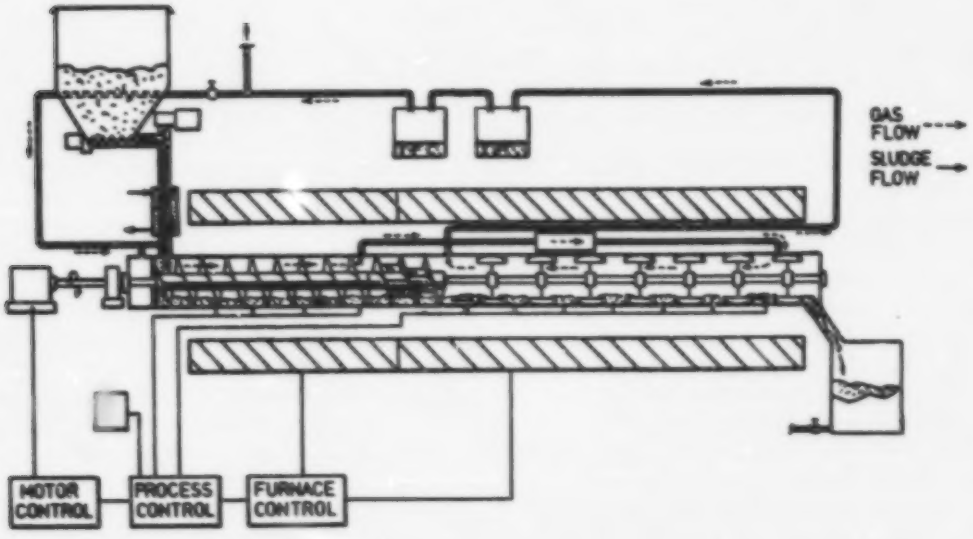

Figure 2. Continuous laboratory equipment

helical gas seal, into a volatilization zone and a char/gas contact zone. The reactor system was designed for dry sludge feed rates of up to 1 kg/h. Solids retention time in the reactor is controlled by varying both the sludge feed rate and reactor inventory. Sludge is fed to the reactor by a calibrated screw conveyor and travels through the reactor by means of the reactor conveyor. Volatilized material is withdrawn in the first zone and can be contacted with the char in both co-current and counter-current modes in the second stage. Product vapours are condensed externally as per the batch reactor system. Inert gas is used to purge the system of oxygen and the operating pressure is less than 1 000 pascals. Prior to collection of experimental data the system was allowed to reach thermal and chemical equilibrium by operating for at least three solid retention times (SRT).

RESULTS

The results will be discussed with respect to both process performance and environmental considerations.

Process Performance - Batch Reactor

Operating test conditions and results from the bulk of the batch experimental program are shown in Table 2. All the data is expressed on a dry sludge basis (corrected for the normal 4-7% moisture present in sludge) and the calorific values are expressed on a total solids basis (not corrected for volatiles). The non-condensable gas (NCG) yield was calculated by difference. Analysis of the NCG, by GC, indicated that it contained roughly 6% methane and 10% carbon monoxide, with the remainder comprising mostly carbon dioxide and nitrogen. The calculated calorific value is approximately 2.0 MJ/kg of NCG.

Optimal test conditions were defined as:

- optimal conversion temperature as determined by differential scanning calorimetry,
- ramping (linear increase with time) to operating temperature at 10°C/minute, and
- continuous nitrogen purge at 7 mL/s.

(i) Different sludges: Sludges from Hamilton, Burlington and Toronto were tested. The average results achieved for these sludges under optimal conditions are shown at the top of Table 2. As can be seen, each of these sludges are amenable to conversion, with oil yields ranging from 20.8 to 24.1% and thermal efficiencies ranging from 77.7 to 83.2%. Throughout this text, thermal efficiency is defined as the energy recovered in the oil, char and NCG as a percentage of that available in the sludge. The average calorific value of the oil generated varied from 33.13 to 37.43 MJ/kg and for the char, from 9.86 to 10.68 MJ/kg. The viscosity of the oils generated appeared to be sludge specific. These variations in performance most likely reflect the variations in sludge quality from site to site. Within the constraints of the experimental program (i.e., batch conditions) the reproducibility within sludge batches was considered acceptable, i.e., generally less than 10% variation in oil yield.

(ii) Effect of Operating Temperature: Operating temperature has a pronounced effect on oil yield. As mentioned, optimal operating temperature was defined using differential scanning calorimetry (DSC). A DSC output for Toronto sludge (Figure 3) indicates two exotherms, one at 300°C and one at 400°C. These exotherms have been attributed to lipid and protein decomposition and subsequent catalysed conversion to hydrocarbons. Oil yield is proportional to the area under these exotherms. It can clearly be seen that a significant reduction in oil yield would be expected by operating at 350°C rather than the optimal of 400-450°C. This was confirmed experimentally using Toronto sludge, (Runs 11, 12, 13 compared to Runs 14, 15, 16) which showed a reduced oil yield from 24.1 to 12.8% with a concomitant decrease in thermal efficiency. However, operating at temperatures above optimal did not appear to significantly affect oil yield or thermal efficiency (Hamilton sludge, Run 22 compared to Runs 1, 20).

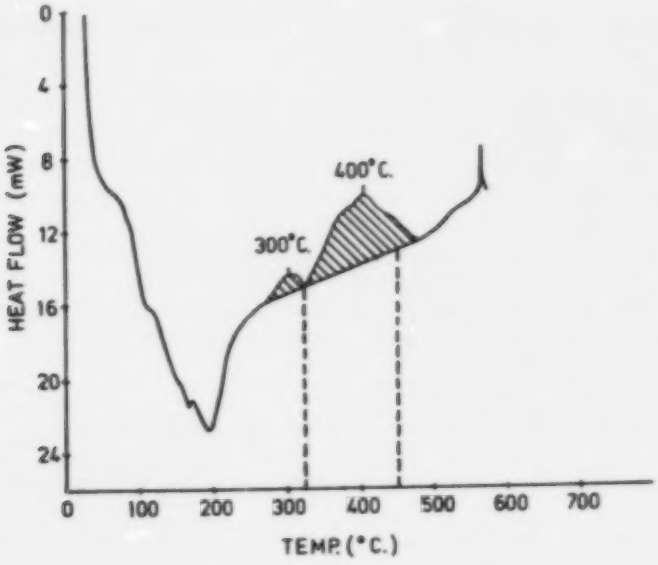

Figure 3. DSC output for Toronto sludge

(iii) Effect of Extended Gas and Gas/Solid Contact Time: A dual reactor system was set up to evaluate whether extended gas and gas/solid contact time had an effect on product yield and quality. The second reactor, identical to the original one, was placed in a vertical position downstream of the first reactor. Vapours from the first reactor were passed through this second reactor prior to condensation. The temperature in the second reactor was maintained at the optimal level.

Three runs were conducted in this mode. The first used an empty reactor (Run 31) to assess the effect of increased gas residence time. The second and third runs (Run 32, 33) were conducted with the second reactor packed with either char or a decarboxylation catalyst (copper).

These two runs were conducted to establish whether increased gas/catalyst contact time had any effects. The data in Table 2 indicate that increased gas/catalyst contact time had a dramatic effect on oil quality. It can be seen that oil viscosity was significantly reduced when extended gas/catalyst contact time was permitted. Under normal conditions

TABLE 2. BATCH REACTOR CONDITIONS AND RESULTS

Run No.	Sludge	Operating Conditions: Temp. (°C)	Operating Conditions: Other Comments	Results: Oil: Yield (%)	Oil: Cal.Value (MJ/kg)	Oil: Viscosity (Centistokes)	Char: Yield (%)	Char: Cal.Value (MJ/kg)	NCG** Yield (%)	Pyrolytic Water Yield (%)	Thermal Efficiency (%)
1,20	Hamilton	400	Optimal	20.8	36.40	Solid*	59.5	9.86	11.6	8.1	81.9
5-10	Burlington	450	Optimal	21.1	37.43	31.1	52.5	10.68	13.2	13.1	77.7
14,15,16	Toronto	450	Optimal	24.1	33.13	60.5	53.7	10.08	13.3	8.8	83.2
11,12,13	Toronto	350	Low temperature	12.8	33.32	Solid	65.6	12.00	10.3	11.2	79.3
22	Hamilton	450	High temperature	22.3	38.87	Solid	54.6	9.39	12.1	11.0	80.4
31	Hamilton	400	Second reactor, empty	19.0	37.49	Solid	60.0	11.07	12.0	9.0	80.1
32	Hamilton	400	Second reactor, char	17.2	38.18	39.5	59.9	11.07	13.0	9.9	77.0
33	Hamilton	400	Second reactor, catalyst	19.0	37.49	31.0	56.8	10.01	14.8	9.4	75.0

*Measured at 20-24°C; solid defined as >214 centistokes

**Non-condensable gas estimated by difference

the oil is solid at room temperature, i.e., viscosity is >200 centistokes. This was reduced to 39 centistokes when char was used as the catalyst and 31 centistokes when a copper catalyst was used. The reduced thermal efficiencies for these runs was probably due to lack of complete product recovery.

(iv) Product Quality: The oils and chars were analysed at CANMET's Energy Research Laboratory in Ottawa. They were characterized using a battery of analyses commonly used to analyse synthetic fuels and the elemental analyses are reported in Table 3. The data in Table 3 indicate that, over the range of process conditions evaluated, the oil and char elemental analyses, with respect to end use, are not significantly different. Thus, while process conditions can affect product yield, it appears that product quality, with respect to elemental analysis, is relatively insensitive to process fluctuations.

Rough elemental balances indicate that 40-50% of the carbon and 30-40% of the hydrogen are classified to the oil. However, only 3-6% of the oxygen, 10-15% of the sulphur and 20-30% of the nitrogen remain in the oil. All of the phosphorous and 75 to 85% of the sulphur remain in the char.

TABLE 3. OIL AND CHAR ELEMENTAL ANALYSES (%)

	Oil					Char				
Run No.	C	H	N	S	O	C	H	N	S	O
20	78.00	10.10	3.99	0.75	6.18	25.45	1.97	2.79	1.39	11.90
9	78.74	10.17	3.45	0.41	6.37	26.02	1.61	3.01	1.16	12.70
15	77.39	9.70	4.95	0.83	6.90	24.53	1.22	2.84	0.74	9.26
22	77.92	10.20	3.99	0.61	6.51	22.53	1.34	2.54	1.52	12.54
31	76.92	10.15	4.11	0.65	6.89	26.53	2.13	2.80	1.31	11.94
32	79.76	10.25	4.19	0.56	5.84	25.97	1.98	2.80	1.34	11.63
33	79.30	10.41	3.49	0.34	5.84	24.22	1.62	2.74	1.50	11.35

The oils were also analysed by sequential elution solvent chromatography to attempt to identify and quantify the constituents present. This analysis indicated that the oils contain approximately 26% saturated aliphatic hydrocarbons, <3% monoaromatics, 1.93% diaromatics and 0.49% polyaromatics. Polar compounds, most likely carboxylic acids, accounted for 28%, whereas 0.9% was basic, pyridine soluble matter, leaving about 39% unaccounted for. A breakdown of the aliphatic hydrocarbons is presented in Table 4.

TABLE 4. ALIPHATIC HYDROCARBON DISTRIBUTION IN OIL

Compound	%
C_{10}	8
C_{10-15}	30
C_{15-16}	6
C_{16-17}	5
C_{17-19}	10
C_{19-20}	10
C_{20-21}	10
$>C_{21}$	21
	100

Process Performance - Continuous Reactor

The continuous reactor system was operated primarily to generate process design information for the full-scale demonstration facility currently under development. Some of this preliminary information is shown in Table 5. Data is presented for sludge feed rates varying from 500 to 1 000 g/h, with the gas/char contactor operated in co-current, counter-current and mixed-flow modes. It was determined that a char inventory of at least 63 g was required to maintain a gas seal between the two reactor zones. Operation with a char inventory below this level resulted in variable and mixed-flow gas routing/contacting.

Comparison of the batch and continuous results indicates similar performance, although there were some differences. Oil yields from the continuous reactor were generally higher than those achieved from the batch system, while yields of NCG and pyrolytic water were lower. The NCG produced from the continuous runs had hydrogen volumetric concentrations of up to 20%, with calculated heating values of up to 20 MJ/kg. Furthermore thermal efficiencies of up to 99% were achieved with the continuous reactor system. These improvements are attributed to the fact that steady-state operating conditions are maintained during continuous reactor runs.

(i) Effect of Operating Temperature: The data generated on the batch reactor indicated that DSC could identify an optimal conversion temperature (for maximum oil yields). This was confirmed by operation of the continuous reactor at above (Run 36) and below (Run 34) the optimal DSC temperature (Run 35). The data support that generated on the batch system, i.e., operation below optimal temperature significantly reduced both oil yield and thermal efficiency, while operation above optimal temperature had little effect on oil yield. As expected, oil viscosity decreased as operating temperature was increased.

(ii) Effect of Gas and Gas/Catalyst Contact Time: A comparison of Runs 37, 38 and 39 illustrates the effect of gas/catalyst contact and gas contact time. Run 37 had ample char inventory to ensure the conveyor seal was operative and the process was truly counter-current, while Run 38 was operated by withdrawing all the evolved gas from the volatilization zone and condensing it directly. Run 39 was operated with the gas seal removed and the gas path routed such that the process was truly co-current. These runs show that extended and uniform contact of the oil precursors produced in the volatilization zone with the char catalyst gives an improved product of reduced viscosity. Thus, optimum counter-current Run 37 produced oil of viscosity 33 centistokes compared to oil of viscosity 73 centistokes for co-current operation (Run 39) and oil of viscosity 110 centistokes for reaction in the volatilization zone only (Run 38).

The quantity of vapour available for reaction to produce oil increases with the feed rate of sludge to the apparatus, while the quantity of available catalyst increases with increase of char inventory. A useful ratio is the char inventory divided by the vapour feed rate expressed as a W/F ratio. The oil viscosity is found to decrease as this ratio increases, this effect being illustrated by comparing the results from Runs 37, 40 and 41. An increase of the ratio from 11 g/g·min (Run 40) to 25 g/g·min (Run 41) is accompanied by a decrease in viscosity from 82 to 34

TABLE 5. CONTINUOUS REACTOR CONDITIONS AND RESULTS

Run No.	Reactor Conditions							Oil			Char		NCG**		Pyrolytic Water	
	Temp. (°C)	Sludge Fd Rt (g/h)	Solids Residence Time (min)	Char inv. (g)	Char inv. / Gas Flow Rate (g/g min)	Gas Seal	Gas Path	Yield (%)	Cal. Value (MJ/kg)	Viscosity* (Centistokes)	Yield (%)	Cal. Value (MJ/kg)	Yield (%)	Cal. Value MJ/kg	Yield	Thermal Efficiency (%)
34	350	688	8	51	12	no	mixed flow	18.53	27.88	160	64.68	8.84	7.28	3.18	3.62	73.85
35	450	728	8	54	11	no	mixed flow	29.71	31.12	72	59.76	8.27	7.57	5.04	2.18	96.80
36	500	707	8	53	10	no	mixed flow	28.16	34.01	34	58.10	7.67	3.61	20.05	3.06	98.04
37	450	758	28	201	33	yes	counter current	24.10	35.53	33	59.50	8.28	6.37	9.68	5.72	93.77
38	450	751	8	56	11	no	1st zone only	24.46	30.06	110	59.47	8.25	9.26	3.72	4.23	83.75
39	450	726	8	55	11	no	co-current	27.96	33.20	73	59.74	8.70	6.67	6.07	5.60	98.91
40	450	911	8	70	11	yes	counter current	26.75	31.04	82	61.11	8.54	4.76	7.20	3.37	92.14
41	450	490	20	88	25	yes	counter current	23.74	35.00	34	57.79	8.12	7.42	10.96	5.88	91.82

*Measured at 38°C (ASTM standard)

**Non-condensable gases

centistokes while an increase to 33 g/g·min (Run 37) is only accompanied by a minor decrease to 33 centistokes. This shows that the ratio should be sufficiently high to ensure adequate and uniform char/gas contact.

These results also show the feasibility of adjusting the distribution of the energy yield among the three products oil, char, and NCG to suit the operating conditions of the apparatus and to maximize the product most required by local markets.

Environmental Considerations

The environmental considerations that need to be addressed before such a process can be commercialized include the following:

(i) process emissions (both aqueous and atmospheric), and
(ii) the impact of product end use (oil, char, NCG and pyrolytic water).

These factors cannot be addressed adequately at this scale of operation, but some valuable information has been generated. Emissions from the process include the NCG and pyrolytic water. It is proposed that at full scale the NCG be combusted with the char to provide the heat needed to drive the process. Based on the limited data available, no new problems are foreseen in this respect. Metal balances indicate that all the metals remain in the char and, therefore, burning should produce an ash very similar to that generated via sludge incineration. It is anticipated that conventional air pollution control technology will be adequate.

The pyrolytic water produced contains approximately 10-15% organic carbon, essentially low molecular weight acids. Based on preliminary analysis, the concentrations of U.S. EPA priority pollutants in this stream are very low. Consequently, this stream also appears suitable for incineration.

The oil has not yet been fully characterized with respect to priority pollutants. Metal analyses do indicate low levels (<5 µg/g) for metals such as Cu, Zn, Cr, V, Pb and Ni. The least valuable end-use for the oil would be to burn it as a fuel oil replacement and no undue environmental concerns are expected in this regard. However, the potential exists to increase end-use value, and upgrading to a transportation fuel will be investigated.

In general, at this point in time, no environmental constraints have been identified which would limit exploitation of the technology.

SUMMARY

Experiments conducted at the WTC, using both batch reactor and continuous reactor systems, have confirmed that Canadian sludges are amenable to conversion to liquid and solid fuels. Oil yields of greater than 25% appear to be readily achievable and thermal efficiencies of greater than 95% easily attained. The data generated on Canadian sludges has in essence confirmed the results generated by the German research and has indicated that this process appears superior to all others referenced in the literature. The most significant fact is that production of straight chain hydrocarbons appears to be unique to this process. The data generated to date supports published claims that net energies of 7-10 MJ/kg of dry sludge should be achievable for the thermal conversion process, provided mechanical dewatering of the sludge is practised(6).

It is estimated that by 1985, 350 000 tonnes of sewage sludge will be incinerated annually in Canada. Thermal conversion of this sludge could produce 700 000 barrels of oil per year, with a market value of at least $21 million.

Environment Canada's long term plans are to demonstrate this technology by construction of a 25 tonne per day facility. The first phase of this demonstration study is currently underway with the development of the pre-engineering data, identification of suitable sites and a thorough assessment of the process economics. This phase of the program is being conducted under contract by Zenon Environmental Inc. and Xytel Energy Ltd.

ACKNOWLEDGEMENT

The authors wish to thank Dr. Henry Sawatzky and his staff at CANMET's Energy Research Laboratory for their technical support and analytical services throughout this study.

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GRANULATION DES BOUES DE LA STATION D'EPURATION DE LA ROCHELLE

par A. DEGUIN*, ing. et Y. JOLY**, ing.

L'épuration des eaux usés urbaines génère d'importantes quantités de boues dont l'élimination cause souvent bien des difficultés aux Municipalités

La station d'épuration de LA ROCHELLE, en France, n'échappe pas à la règle et jusqu'ici les boues produites devaient être transportées jusqu'à une décharge publique.

D'autre part, à cause de la vocation particulière du port de commerce de LA ROCHELLE recevant de grandes quantités de bois exotiques, 5 à 6 000 tonnes/an de déchets et d'écorces de bois devaient aussi être transportés et éliminés.

Afin de résoudre le problème des boues, le Syndicat intercummunal de la région de LA ROCHELLE à confié au groupe SAUR (Société d'Aménagement Urbain et Rural) la responsabilité d'étudier, construire et opérer une installation de séchage et valorisation de ces boues. Le contrôle de ce mandat a été effectué par la direction départementale du Ministère de l'Equipement, agissant ainsi à titre de Maître d'Oeuvre.

SAUR a répondu à cette demande en résolvant ensemble les deux problèmes, celui des boues et celui des écorces. Ainsi, l'utilisation de l'énergie provenant de la combustion des déchets de bois permet de conditionner thermiquement et sécher les boues de la station d'épuration.

Cette installation de granulation a été construite en 1983 et fonctionne à présent avec succès depuis près d'un an.

I - STATION D'EPURATION DE LA ROCHELLE

Prévue pour 140 000 équivalents habitants, cette station ne reçoit actuellement que les 3/4 de sa charge. Elle traite donc environ:

- 20 000 m^3/d soit 5.5 MGUSD d'eaux usées et un peu plus de 6 tonnes de DBO_5 par jour.

La file de traitement comprend:

- un prétraitement avec dégrillage, dessablage et dégraissage,
- une clarification primaire,
- un traitement biologique par boues activées à moyenne charge avec insufflation d'air par grilles,
- une clarification secondaire dans un décanteur lamellaire,
- une deshydratation par filtration sous vide des boues digérées en anaérobiose,
- une deshydratation par filtre-presse des boues fraîches épaissies.

Le volume de boues séchées à éliminer chaque jour est d'environ 35m^3 (9200 GUS). Ces boues sont conditionnées à la chaux et au chlorure ferrique. Leur siccité est en moyenne de 18% à la sortie du filtre sous-vide et de 35% après le filtre-presse.

La production théorique annuelle de boues serait de 12500 t, à la capacité nominale de la station.

Ces données ont servi de base au design de l'installation de granulation des boues.

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2: INSTALLATION DE VALORISATION DES BOUES PAR VOIE THERMIQUE

Comme on vient de le voir, l'originalité du choix de la source d'énergie nécessaire à la granulation des boues constitue la solution à un double problème:

- l'enlèvement des boues de la station d'épuration,
- l'élimination des écorces de bois exotiques.

Pour ce faire, il a fallu réaliser deux installations:

- l'une, pour broyer les écorces en vue d'homogénéiser le combustible du four de granulation,
- l'autre, pour réaliser la granulation des boues.

2.1 Broyage et stockage des écorces

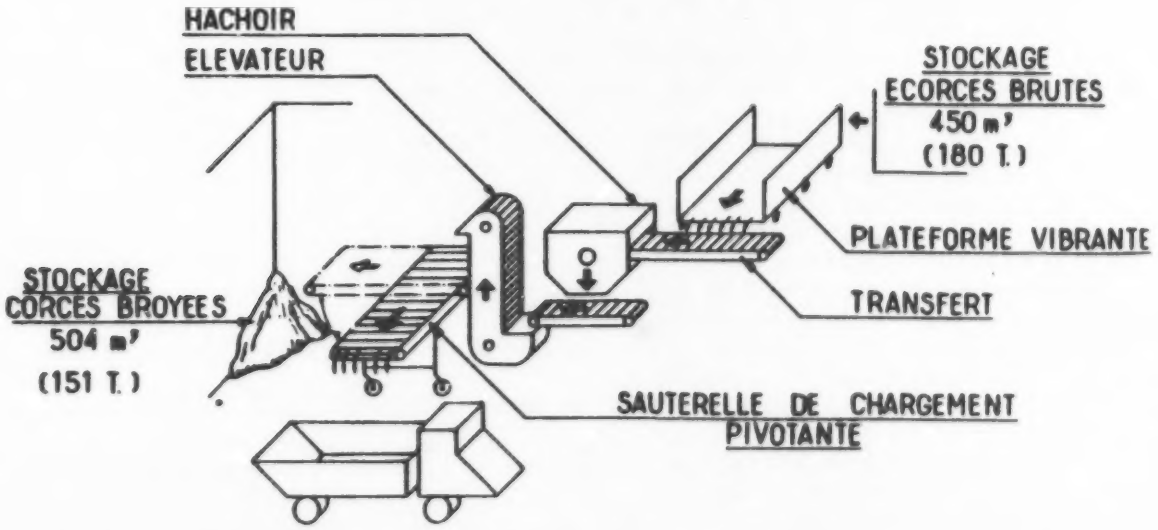
La consommation annuelle d'écorces pour le sèchage des boues est estimée à 2 500-3 000 tonnes.

L'unité de broyage et de stockage des écorces est située sur les dépendances du Port, distantes de 2 km environ des installations de sèchage de boues.

Les écorces, ramassées sur les terre-pleins portuaires, sont stockées sur une aire de 325m^2 permettant l'accumulation d'environ 180 tonnes de matière. Elles sont reprises de cette aire par un engin mécanique pour les véhiculer jusqu'à une plate-forme vibrante (voir schéma no.I) destinée à les orienter dans le sens de la longueur.

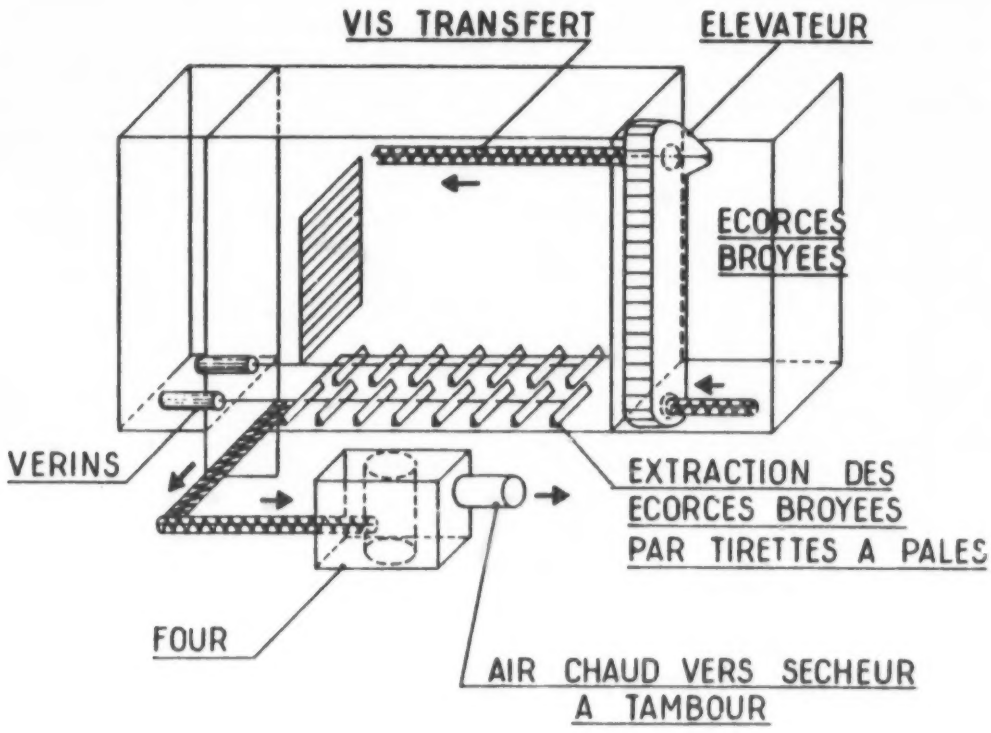
Elles sont ainsi dirigées vers un hâchoir à couteaux qui les déchiquette au débit de 2t/h. Les petits morceaux d'écorces, de 3 à 4 cm, sont alors repris par des élévateurs en direction d'un convoyeur à bande les orientant soit vers une aire de stockage (capacité 500 m3/150 tonnes), soit vers un camion-benne. Celui-ci assure le transport des écorces broyées jusqu'au silo de la station de granulation.

Ce silo, de 130 m3 (40 tonnes) est accolé au bâtiment de granulation. Il permet une autonomie de plus d'une semaine à raison de 10 heures de fonctionnement par jour.



1 - SCHEMA DE L'INSTALLATION DE BROYAGE DES ECORCES

Le déchargement des écorces provenant de l'unité de broyage se fait par déversement de la remorque sur un doseur vibrant qui les oriente sur un élévateur à godets, lequel remonte les écorces au sommet du silo de stockage (voir schéma no. 2).



2 - SCHEMA DU DISPOSITIF DE CHAUFFAGE DES BOUES

L'extration des écorces du silo s'effectue par des râcleurs à échelle coulissant sur le sol et manoeuvrés par vérins hydrauliques. Ces râcleurs guident le matériau sur une vis transporteuse placée au sol. Repris ainsi par deux vis successives inclinées, le combustible aboutit alors dans la chambre de combustion du four.

2.2 Installation de granulation

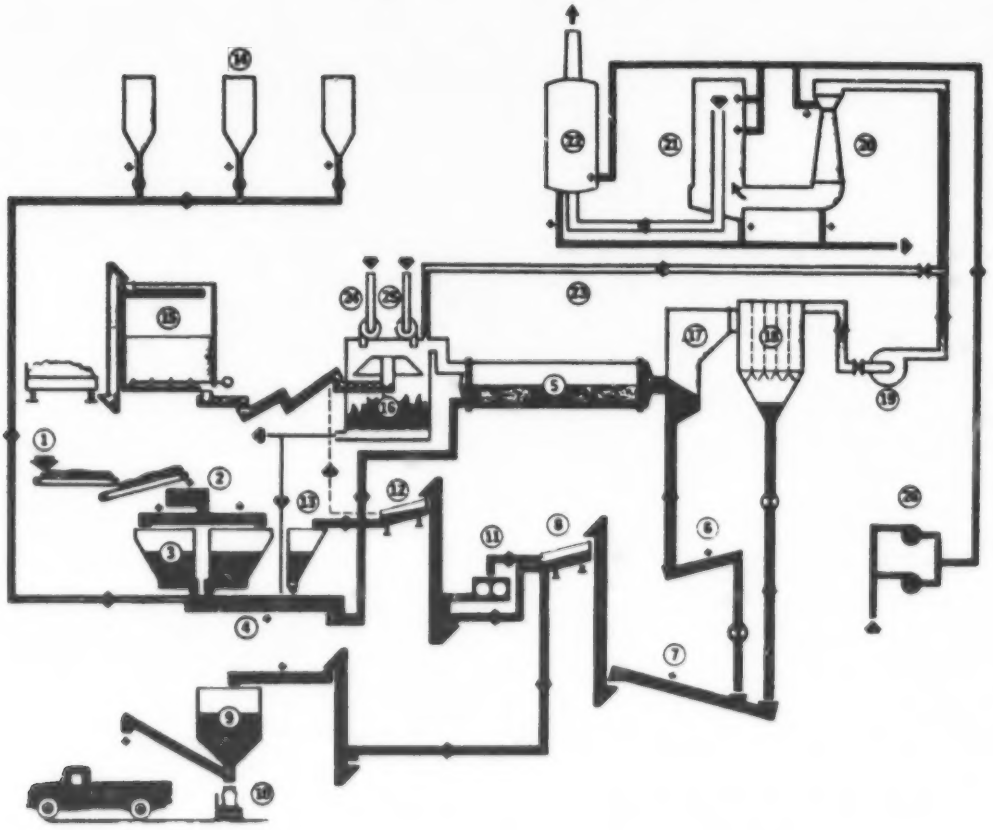
Le principe du procédé consiste à fabriquer par sèchage à haute température de petits granulés de boues de 2 à 4 mm de diamètre. La formation de ces granulés se produit par initiation à partir de granulés plus fins recyclés dans l'installation. Cette granulation se fait dans un tambour sècheur rotatif à pales internes.

Lors de la première mise en service, l'initiateur peut être de la sciure de bois autour de laquelle se formera un enrobage de boue sèchée. Cet enrobage grossira au fur et à mesure du transit des grains dans le tambour.

Les granulés sortant du tambour sont tamisés pour ne produire que des grains de taille sélectionnée. Les granulés trop gros sont broyés, mélangés aux grains trop fins et recyclés.

L'air chaud insufflé du four vers le sècheur à tambour est dépoussièré, lavé et désodorisé avant rejet à l'atmosphère (voir schéma no. 3)

3 - SCHEMA DE FONCTIONNEMENT



1 Arrivée des boues déshydratées
2 Emotteur
3 Silos de stockage des boues humides
4 Vis mélangeuse
5 Tambour sècheur
6 Vis de transfert des granulés lourds
7 Vis de transfert des granulés
8 Tamis vibrant
9 Stockage du produit fini
10 Ensachage ou chargement des granulés
11 Broyeur
12 Tamis vibrant
13 Stockage ensemencement
14 Silos pour adjuvants
15 Stickage écorces broyées
16 Four
17 Chambre de détente
18 Multicyclone
19 Ventilateur principal
20 Laveur de fumées type tuyère
21 Tour de lavage
22 Tour de désodorisation
23 Air de recirculation
24 Air primaire
25 Air secondaire
26 Pompes à eau

2.2.1 Production de chaleur. Le four est alimenté en écorces broyées par une vis. Celle-ci est disconnectée du stockage des écorces pour des raisons évidentes de sécurité. Elle passe sous le four et débouche à la base du puits d'alimentation du foyer.

Ce dernier est à grille tournante. Il a la forme d'un tronc de cône dont la grande surface constitue la base.

L'installation de combustion produit le gaz chaud destiné à sècher les boues. Elle consomme environ 500 kg de bois à l'heure.

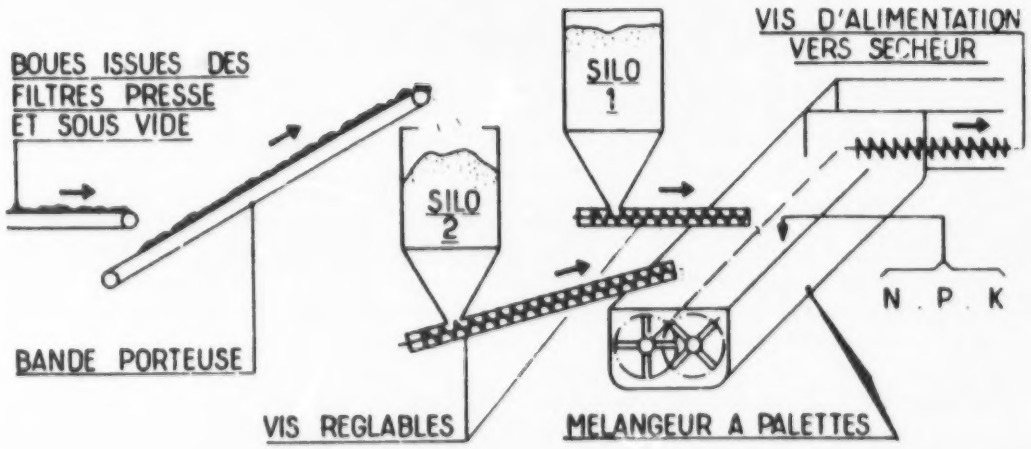
Les cendres sont râclées et recueillies dans une vis de transport qui peut les recycler dans la boue à sècher. L'apport de potasse ainsi obtenu est un facteur non négligeable pour la valorisation agricole des boues. Toutefois, le machefer résultant de la combustion des écorces limite ce recyclage.

La quantité de cendres est de l'ordre de 0,5 à 1 % du poids de bois brûlé.

Le rendement du four est amélioré par recyclage d'une partie des gaz chauds réintroduits en tête de chaudière après dépoussièrage. La puissance évaporatoire de l'installation peut ainsi représenter 2 tonnes d'eau à l'heure.

2.2.2 Sèchage des boues. Après leur déshydratation, les boues de la station d'épuration sont véhiculées par bande transporteuse vers deux silos. Avant d'y être stockées, les boues sont émottées.

Elles sont extraites ensuite de ces silos et transportées par vis jusqu'à un mélangeur à palettes (voir schéma no. 4). Ce dernier reçoit des additifs fertilisants du sol, éventuellement choisis pour l'obtention d'une composition finale spécifique. Il reçoit aussi les cendres du four et de l'eau permettant, le cas échéant, la correction du point d'humidité du mélange.



4 - SCHEMA DU MELANGEUR DE BOUES

Ce mélange est introduit par vis dans le tambour sècheur où se produit la granulation.

Le sècheur est un grand cylindre métallique calorifugé, de 1,90 m de diamètre et 8,50 m de long. Sa rotation à 4 ou 5 tours par minute est assurée par un moteur électrique de 10 HP.

L'appareil est divisé en sections avec un tunnel central, permettant l'évacuation du produit sec au fur et à mesure de sa deshydratation, tout en retenant les particules encore trop humides.

Seules les particules devenues assez légères sont entraînées par le courant d'air chaud issu du four.

Cet air, aspiré par un ventilateur pénètre dans le sècheur à environ 700°C-800°C (1300-1500°F).

En sortie de tambour, dans la chambre de détente, les granulés secs sont séparés gravitairement des fumées. Celles-ci s'élèvent en direction du multicyclone.

2.2.3 Tri et conditionnement des granulés. Les granulés recueillis au bas du séparateur primaire et les fines piégées dans le multicyclone (voir schémas no. 3 et 5) sont transportés par vis jusqu'aux tamis vibrants.

Les granulés de taille commerciale (3 à 4 mm de diamètre) sont repris par un élévateur à godet.

Les refus sont concassés dans un broyeur à marteaux et réunis avec les fines pour être transportés, toujours par un élévateur à godet, vers le silo de stockage des éléments d'ensemencement. Ceux-ci sont orientés en fonction des besoins, sur le mélangeur à palettes.

Les granulés de taille commerciale sont refroidis à l'air avant stockage en silo. Cet air est recyclé sur le four.

Selon les besoins, on peut réaliser un ensachage ou un transport par vis pour évacuation en vrac.

2.2.4 Traitement des fumées. Les gaz de combustion, mélangés à la vapeur d'eau provenant de la deshydratation des boues, sont traités par:

- séparation des poussières par une batterie de cyclones,
- lavage des fumées par aspersion d'eau à l'aide de tuyères,
- désodorisation sur tour.

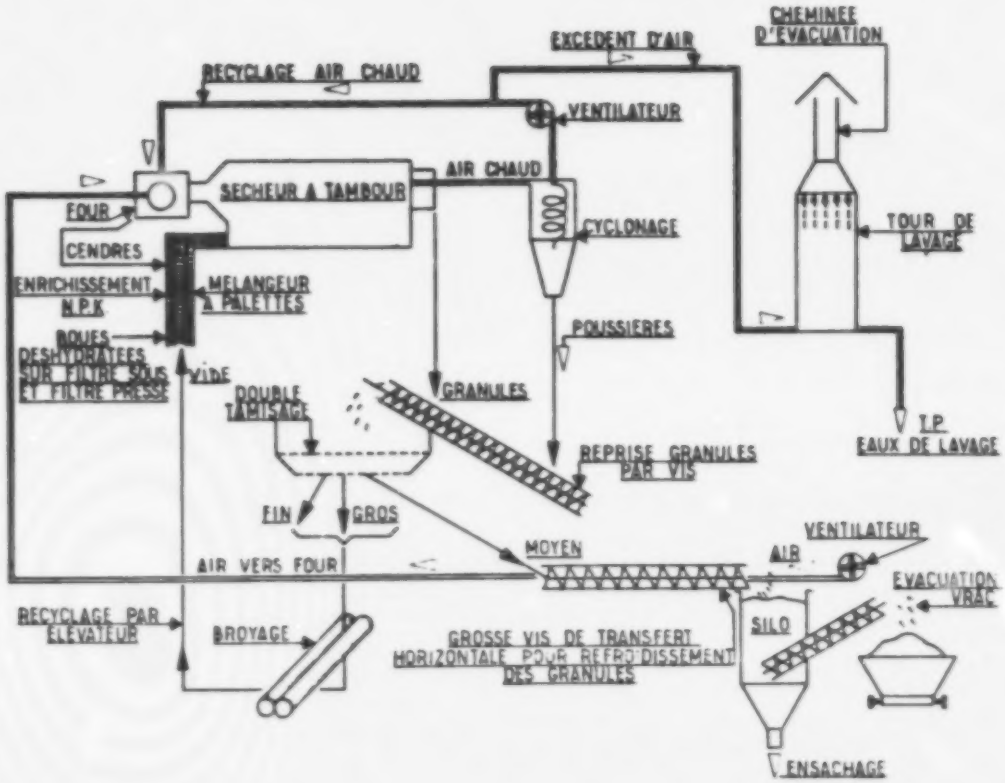
En sortie du multicyclone, les fumées ne doivent pas contenir plus de 200 mg de poussières par mètre cube dont les trois quarts devront être éliminés sur les différents étages ultérieurs de lavage.

Les fumées débarassées des poussières les plus lourdes sont aspirées par un ventilateur qui les refoule vers la base de la tour de lavage.

L'eau d'aspertion est prélevée par pompage en sortie des décanteurs secondaires de la station d'épuration.

La température en sortie de la tour de lavage chute à environ 50°C (120°F).

Les fumées sont ensuite introduites à la base d'une tour de désodorisation garnie d'anneaux Raschig au sommet de laquelle on pulvérise un agent chimique oxydant (hypochlorite de sodium) ou acide.



5 - SCHEMA DU DISPOSITIF DE SECHAGE ET DE GRANULATION DES BOUES

3. RESULTATS D'OPERATION DE L'INSTALLATION

Après quelques mois nécessaires à la mise au point du fonctionnement, l'installation fonctionne depuis le début de l'année 1984. L'exploitation est assurée par une entreprise privée (SAUR). Il est encore trop tôt pour dégager un bilan précis de cette opération mais il peut d'ores et déjà être fait état des premiers résultats obtenus.

Le tableau no. 1 révèle qu'après les deux premiers mois de démarrage de l'exploitation des installations, la production réelle de granulés rejoint pratiquement celle attendue en théorie, avec un rendement de 94 %.

TABLEAU N° 1 : PRODUCTION DE GRANULES

ANNEE 1984	Boues filtre presse (T)	M.S. filtre presse (T)	Boues digérées (T)	M.S. boues digérées (T)	Total M.S. (T)	Granulés théoriques (T) (à 90 % M.S.)	Granulés réels (T)	Rendt %
Janvier	289	96,5	128	28	123,5	137	57	42
Février	435	161	46	10	171	190	99,5	52
Mars	429	142	96	25	167	185,5	175	94
Avril	377,5	136,5	75	16	152,5	169,5	160	94
TOTAUX	1 530,5	535	345	79	614	682	491,5	72

Le tableau no. 2 rend compte du tonnage d'écorces broyées pour les besoins de consommation de l'unité de granulation.

En quatre mois on a brûlé 579 tonnes de déchets de bois, ce qui conduirait dans ces conditions à une consommation annuelle d'environ 1750 tonnes. En fait la production prévisible de granulés pour l'année 1984 devrait atteindre 2000 tonnes, compte tenu des faibles productions de janvier et de février.

Lorsque la station d'épuration traitera sa capacité nominale d'eaux usées, la production théorique de granulés devrait être de 3 500 à 3 600 tonnes par an.

TABLEAU N° 2 : BROYAGE DES ECORCES

ANNEE 1984	Ecorces broyées (tonnes)	Horaire de broyage (heures)	Capacité (T/h)
Janvier	140	103	1,36
Février	158	104,5	1,51
Mars	146	90	1,62
Avril	135	82,5	1,64
TOTAL	579	380	1,52

3.1 Caractéristiques du produit fini

Les études prévisionnelles établissaient la qualité des granulés aux niveaux suivants:

- humidité maximum ..... 12 %
- teneur en matières organiques minimim ..... 45 %
- teneur en azote organique minimum..... 3 %
- teneur en métaux lourds telle que le produit soit conforme à son emploi au titre de fumure des sols.
- poussières.....- inférieures à 10 %

Les premiers résultats obtenus donnent un produit respectant le cahier des charges avec une densité apparente voisine de 0,55 pour une production ne correspondant qu'à 60 % seulement de la capacité nominale.

La composition moyenne du produit fini est la suivante:

- siccité ..... 92 à 93 %
- teneur en matière organique ..... 48 à 55 %
- teneurs par rapport aux matières sèches:
  - . azote total .....2,7 à 2,8 %
  - . azote ammoniacal .....0,4 à 0,45 %
  - . azote organique .....2,3 à 2,4 %
  - . phosphates totaux (P_2O_5).....4 à 5 %
  - . phosphates solubles (P_2O_5).....3,2 à 3,5 %
  - . potassium (K_2O) .....0,2 %
  - . Calcium (CaO) .....15 à 19 %
  - . métaux lourds inférieurs à la norme

Nous observons un léger déficit en azote organique résultant d'un exédent de calcium ajouté à la boue lors de la deshydratation mécanique dont l'obtention d'un produit correct oblige à l'application de taux élevés de chaux.

Un projet de deshydratation par centrifugation des boues avant sèchage et granulation en remplacement des filtres actuellement utilisés, devrait assurer une amélioration sensible des résultats.

En outre, l'installation est équipée de la possibilité d'enrichir la boue en éléments fertilisants si la Société chargée de commercialiser le produit fini en émet le désir.

Ainsi des silos, avec dispositifs d'injection de produits azoté, phosphoré ou contenant du potassium, permettent-ils d'adjuster à la demande les doses en N, P, K de la boue, au niveau des mélangeurs placés en amont de l'admission dans le tambour-sècheur.

Les granulés sont livrables soit en vrac soit en sacs polyéthylène de 25, 33 ou 50 kg.

Leur épandage avec les machines agricoles classiquement utilisées pour fumer les sols avec des produits secs peut se faire sans problèmes de même que leur stockage.

3.2 Coûts. Conformément aux dispositions du règlement du concours, SAUR assure l'exploitation de l'installation pour une durée de 10 ans en garantissant lèlement la commercialisation des granulés.

Le coût de construction de l'ensemble des installations (broyage et granulation) est de 15 millions de francs TTC (environ 2,3M$), subventionné en partie par l'Etat et l'ANDRED.

Les recettes procurées par la vente des granulés auxquelles il convient d'ajouter les économies réalisées sur le transport et déchets (boues et écorces) doivent progressivement amortir les dépenses d'investissement en couvrant en tout temps les frais d'exploitation.

Ceux-ci proviennent principalement des coûts de main-d'oeuvre (4 employés), d'entretien et de renouvellement du matériel. Les coûts d'électricité ne représentent que 50 % des coûts de main-d'oeuvre.

4. CONCLUSION

L'installation de valorisation des boues de la station d'épuration de LA ROCHELLE, permet de supprimer la nuisance et le coût d'une évacuation en décharge tout en assurant une gestion équilibrée par la vente du produit fertilisant obtenu, ce qui constitue une solution satisfaisante du problème posé par ces boues.

Mais en plus, cette installation allie l'économie à l'efficacité en utilisant comme source d'énergie des écorces, éliminant du même coup une nuisance supplémentaire de l'environnement local.

La preuve est ainsi faite que la valorisation des boues par voie thermique peut être viable en dépit du coût croissant des sources d'énergie habituelles.

ALTERNATE NITROGEN REMOVAL TECHNOLOGY: PILOT STUDIES APPLIED TO LOCAL PROBLEMS

Christopher Biemiller
Suffolk County Department of Public Works

ABSTRACT

Faced with self determined upgrading requirements to remove nitrogen from wastewater discharges of existing treatment works and of non-sewered commercial developments, two separate departments within a County structure undertook pilot studies, demonstrated these as practical, and are now applying the results of these efforts to full scale projects. By their own initiative, these two County departments brought low cost/no cost solutions to self determined environmental problems.

Summary

As a result of decisions made to protect the groundwater supply reservoir of Suffolk County (New York) from nitrate contamination by subsurface wastewater disposal, the majority of existing secondary treatment plants are under legally binding consent orders to provide the advanced treatment necessary to conform to a 10 mg/l total nitrogen effluent discharge limitation. Implementation of the upgrading requirements had been deffered pending studies undertaken by the two County agencies most intimately involved with wastewater disposal, the departments of Health and Public Works.

Health's studies were of passive removal techniques modifying the cesspool and leaching field disposal method most commonly used for residences and small commercial systems in the rural non-sewered areas of the County. As developped the system now being specified by Health involves fixed growth nitrification on aerobic gravel substrate followed by sulfur/limestone denitrification. Figure A is a schematic of the system as currently installed at over fifty (50) sites treating flows ranging from 30 to 900 l/s.

Public Works' studies were of "no cost/low cost" modifications of their existing extended aeration activated sludge facilities, designed originally for secondary treatment of BOD and suspended solids. The study reported here is a suspended growth single sludge nitrification denitrification modification of the extended aeration structure. No additional reactor tankage or clarifiers are required. The acceptable performance of this modification, particularly when balanced against its "no cost" price, has caused Public Works to propose this upgrading technology for five other facilities ranging in size from 5 m^3/s to 22 m^3/s, treating a total of 62 m^3/s. Figure B is a schematic of this modification.

Together the efforts of these two departments are being directly applied to current needs. A problem was percieved and a solution, although perhaps not entirely understood or tested, is provided as a technical and operational reality.

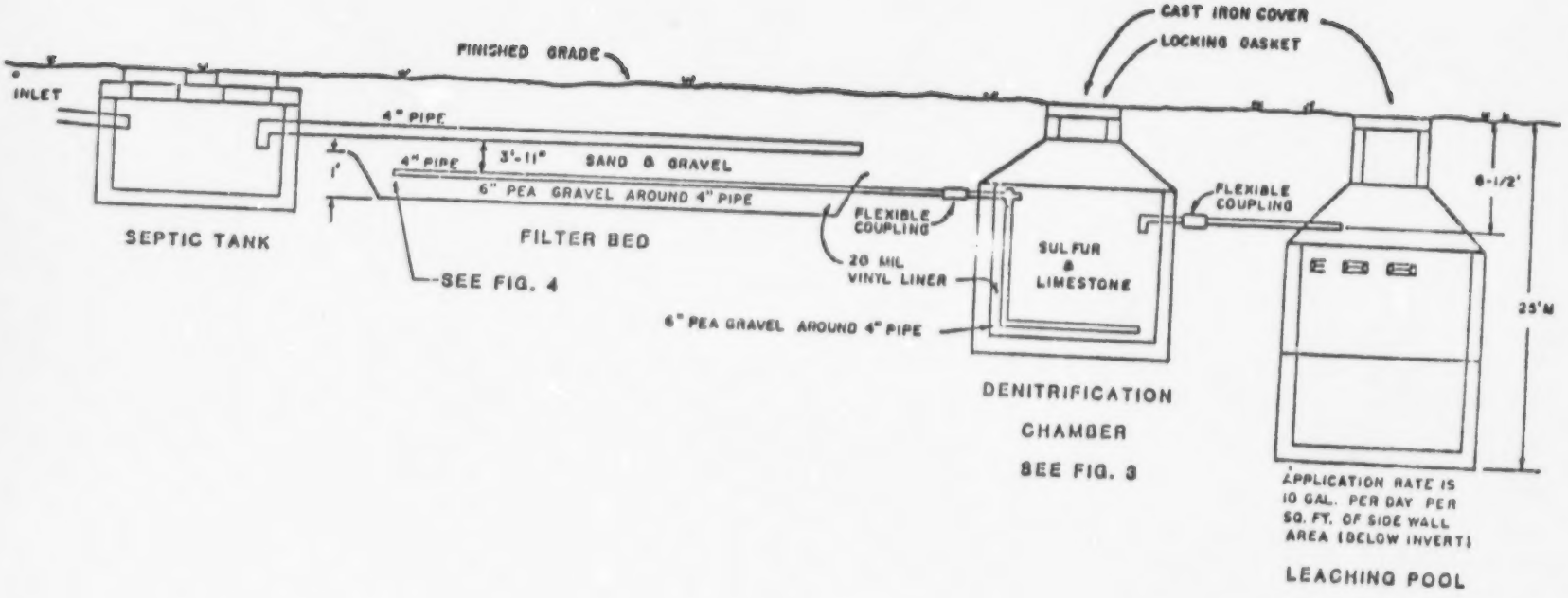


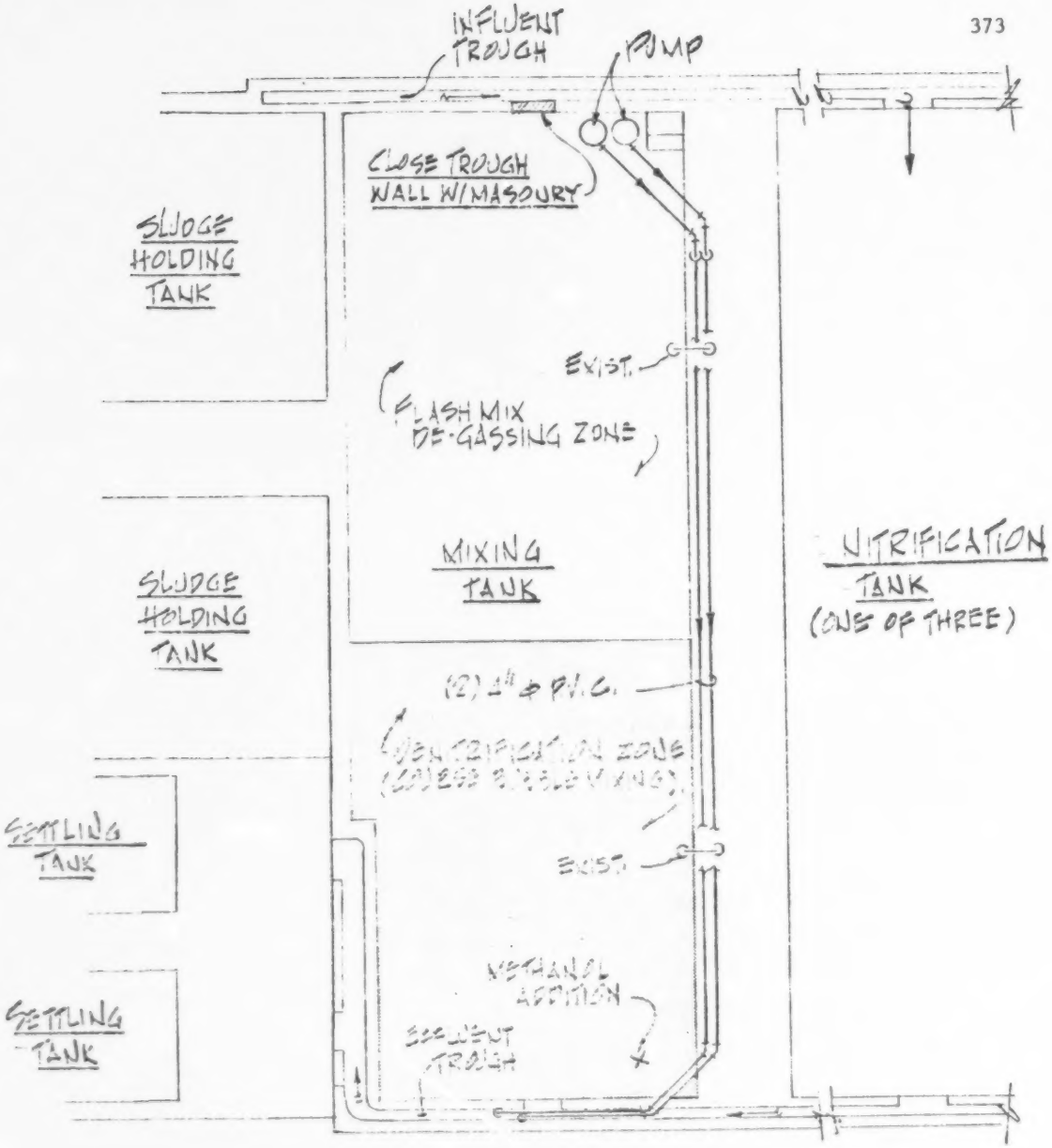
FIGURE A

GENERAL PROFILE

DENITRIFICATION SEWAGE TREATMENT SYSTEM

GRAVITY FLOW

NO SCALE



MODIFICATION OF PROCESS AT PARKLAND III S.T.P.

REF: NELSON & PAPE DWG. SH 9 OF 12

FIGURE B

DATE: 3-83
DR. RR. APP'D B.W.

POOR

QUALITY

ORIGINAL